CONVERSION OF COAL REJECTS TO LIQUID FUELS VIA INDIRECT LIQUEFACTION

FINAL REPORT

PREPARED BY:

WESTAR MINING LTD.
ESSO RESOURCES CANADA LTD.

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EXECUTIVE SUMMARY

Objective

The aim of this $292,000 study was to check the technology and economics of converting the coal refuse at the Elkview Preparation Plant of Westar Mining Ltd. to premium gasoline via an indirect process.

Award

The Department of Supply and Services (DSS) signed a 50:50 Cost Sharing Contract with Westar Mining Ltd. in June, 1984. Esso Resources Canada Limited agreed with Westar to share equally in the venture. The study is part of the Energy Conversion Program. Dr. E. Furimsky of Canmet provided technical supervision.

Previous Work

In previous Government funded studies with Westcoast Transmission and PetroCanada, direct liquefaction of coal did not appear economic. As an extension, a small team examined conversion potential using low value feedstock, still rich in carbon (34 - 52%) but high in ash (28 - 51%), at the rate of 9,600 dry tonnes/sd feed. With this unusual feedstock an indirect process appeared most suitable. The economic results of this preliminary examination were considered to be sufficiently attractive to further study the process.

Process Description

There are four (4) main phases in the overall process:

1. Refuse Preparation - grinding, drying and storage;
2. Gasification - using Koppers Totzek (KT) or Texaco processes;
3. Methanol Production - by the proven ICI process; and
4. Methanol to Gasoline - by the Mobil process (MTG).

Ancilliary oxygen plant, acid gas removal plant, sulphur plant, utilities systems, administration and maintenance buildings are also required. These are provided on site across the river from the Elkview Plant, sized for a feed rate of 4,800 tonnes/sd.

Results

Exxon Research and Engineering Company (ERE) evaluated the technical and cost aspects of the liquefaction plant. Westar Engineering Ltd. designed the refuse handling arrangements.

Since the previous study, experience with the newer Texaco gasifier has added to the knowledge base and raised confidence in this system. However, the base case, as before, is costed using the KT process. A Texaco scheme is added as an alternative.

Feeding 4,800 dry tonnes per stream day of Elkview Plant refuse, the following production rates are achieved:

Unleaded Gasoline	1281 m^3/sd (8,063 Bpsd)
Butane	182 m^3/sd (1,147 Bpsd)
Propane	99 m^3/sd (621 Bpsd)

Economics

Some assumptions were necessary, as follows:

Feedstock (refuse) Value	$6.00 per dry tonne
Selling Price of Gasoline - 1985	$.22/litre (C$ 35/Bbl.)
Selling Price of Gasoline - 2000	$.63/litre (C$100/Bbl.)

Approximately 30% of gasoline revenue normally received by Governments as gasoline taxes flows back to the project as additional revenue.

Estimated capital required:	1.51 Billion (2 Qtr. 1984).
DCF rate of return, base case:	4.3%
DCF rate of return, 20% extra revenue:	7.5%
DCF rate of return, base and Texaco gasifier:	6.5%

Subsidy

To return 15% DCF, a subsidy of $2.35 Billion (1984$) is required over the life of the project - equal to about the price of a barrel of crude oil per barrel of product.

Markets

The comparatively small production of 1,300 m^3/day or 8,000 barrels per day of gasoline can be absorbed in the market which is projected to be growing slowly. Butane and propane are expected to sell at premium prices in the North West U.S.

Alternative Process Configurations

The possibility of using two alternative processes to produce chemical synthesis gas or methanol were examined. The returns are again unattractive. To achieve a 15% DCF at today's prices requires a subsidy of $1.6 Billion (1984$) over the project life.

Recommendations

Due to poor economics at present prices, it is recommended that no further work be done until there is firm evidence that crude oil and produce prices are escalating dramatically.

1.0 INTRODUCTION

1.1 Project Background

Located in the southeast corner of British Columbia, the Elk Valley possesses one of the richest deposits of coal in Canada. Several major coal mining companies extract, wash and export the abundant energy resources of this area. The major coal mines are:

Westar Mining Ltd.	Balmer Mine - Sparwood
	Greenhills Mine - Elkford
Shell Resources Ltd.	Line Creek Coal Mine - Sparwood
Esso Resources Canada Ltd.	Byron Creek Mine - Corbin
Fording Coal Ltd.	Fording Coal Mine - Elkford

The Balmer coal mine of Westar Mining Ltd., has been in operation for about 15 years and coal rejects produced at its Elkview Coal Wash Plant are stored in coarse reject stockpiles and in fines reject lagoons. Petro Canada performed a pre-feasibility study to examine the potential for conversion of these coal rejects to liquid fuel via indirect liquefaction in 1982-3.

The study results are contained in a report titled "Western Canada (B.C.) Coal Liquefaction Study - Phase IA".

Based on the results of the Phase IA report a consortium consisting of Esso Resources Canada Ltd., Energy Mines and Resources Canada and Westar Mining Ltd. agreed to reconsider the scope and direction of the indirect coal liquefaction opportunity in S.E. British Columbia.

1.2 Terms of Reference

The Phase IA Study was completed in early 1983. In November 1983 the Federal Government, through the Department of Supply and Services, issued a request for proposals to study the conversion of coal to liquids under the 50/50 cost shared Energy Conversion Programme. Westar Mining Ltd. in association with Esso Resources were awarded a contract to perform the study which seeks to review the previous report and calculate the economics at half scale; i.e. the processing of 4,800 dry tonnes per stream day of coal rejects instead of 9,600 dry tonnes/stream day. The technical officer is Dr. Ed Furimsky from Canmet.

In this contract, specific tasks are allotted as follows:

Task 1: Critical Assessment of Existing Processes Used for the Indirect Liquefaction

- o Examine gasifier requirements, i.e. re-evaluate the Texaco and Koppers-Totzek technologies.
- o Select one (either Texaco or Koppers-Totzek) for this study.
- o Review plant design for 4,800 dry tonnes per stream day feed rate, using modular design.
- o Determine interfaces with refuse preparation plant, etc.
- o Study CO shift requirements.
- o Study methanol plant.
- o Confirm Methanol to Gasoline (MTG) costs and performance with Mobil.
- o Check material balance and energy balance.
- o Check water requirements and availability.
- o Check power requirements, power supply and re-use of water heat. (Net exothermic process).
- o Fix inventory requirements for feed and product.
- o Check necessity for sulphur removal unit.
- o Prepare or modify plant and circuit drawings and cost estimates.

Task 2: Review and Assessment of Coal Rejects as a Feedstock for Indirect Liquefaction

- o Review refuse availability, handling and preparation system.
- o Jointly select location of refuse preparation and chemical plant.
- o Determine flow from Elkview cleaning plant to chemical plant.
- o Review refuse upgrading process.
- o Prepare flow diagram and materials balance.
- o Prepare capital and operating cost estimates.
- o Consider disposal of new refuse from the chemical plant.

Task 3: Economic Assessment of the Proposed Route

- o Prepare capital and operating costs for plant following the reviews in Task 2.
- o Evaluate commercial arrangements of licensors.
- o Prepare sensitivity analysis.
- o Assess the market impact and establish revenue streams.
- o Identify possible environmental conflicts.

Task 4: Report Preparation and Go-No-Go Decision

- o Meeting of partners to allocate work and decide on method of presentation.
- o Preparation of narrative.
- o Preparation of technical drawings.
- o Preparation of Economic analysis.
- o Derivation of conclusions.
- o Decision to proceed to the gasifier test, or to terminate.
- o If the decision is to terminate, the draft final report will be finalised.
- o If the decision is to proceed, to go to Task 5.

Task 5: Preparation of Refuse Samples by Westar Staff

- o Discussion with Licensor to arrive at composition of *three typical samples from the four refuse streams:

  Heavy media vessel +9.5 mm
  Heavy media cyclone 9.5 mm x 0.6 mm
  Water only cyclones 0.6 mm x 0.25 mm
  Flotation tailings minus 0.25 mm.

- o Strict supervision to collect and blend the samples.
- o Packing the samples.
- o Obtain quotations for transport to testing facility.
- o Follow-up on safe arrival at test facility.

* the number of samples is subject to change, but the projected cost of the tests must not be exceeded.

Task 6: Gasifier Tests

- o Licensor will conduct the tests.
- o One technical man from the consortium (or a consultant) will witness the tests, expected to take 2 or 3 days.
- o Discussion of preliminary results with licensor.
- o Submission of full report by licensor accompanied by guarantee of performance, with estimates of capital and operating costs.

Task 7: Review of Tests by Technical Committee of Consortium

- o Meeting to discuss the results.
- o Inclusion of results into the economic study.
- o Completion of draft final report.
- o Printing of final report (15 copies).

1.3 Scope of Work

The scope of work for the indirect liquefaction study can be summarized in four phases:

Phase I - Review of Phase IA report and update on gasification process selection decision.

Phase II - Detailed definition of coal preparation plant and site selection.

Phase III - Update of costs and economics for a half-scale plant from the Phase IA study.

Phase IV - Execution of a gasification demonstration test run on the preferred technology.

Westar Mining Ltd. contracted with Westar Engineering Ltd. for overall project management as well as engineering work on site selection and refuse preparation plant design. Esso Resources Canada Ltd. (ERCL) were responsible for reviewing the gasifier process selection decision as well as updating costs and economics for the half scale plant. ERCL enlisted the help of Exxon Research and Engineering to accomplish these tasks.

The principal participants in these respective organizations are as follows:

David M. Parkes - Westar Engineering - Project Manager
Pyara S. Lotay - Westar Engineering - Principal Engineer
Lindsay D. Dowdle - Esso Resources - Principal Engineer
David C. Bandy - Esso Resources - Financial Analysis
Exxon Engineering - Petroleum and Synthetic Fuels Department -Gasifier Technology
Exxon Engineering - Project Management Department - Cost Estimating

2.0 FEEDSTOCK PROPERTIES

2.1 General

The coal rejects produced by Westar Mining Ltd.'s Balmer mine Elkview preparation plant at Sparwood, British Columbia were selected as feedstock for the coal to gasoline conversion complex.

The Elkview plant produces four reject streams and for this study phase, three of these streams are utilized to provide raw material feedstock to the gasification plant. A survey of the present coal reject disposal areas indicates the following availability of feedstock material:

Fine Refuse

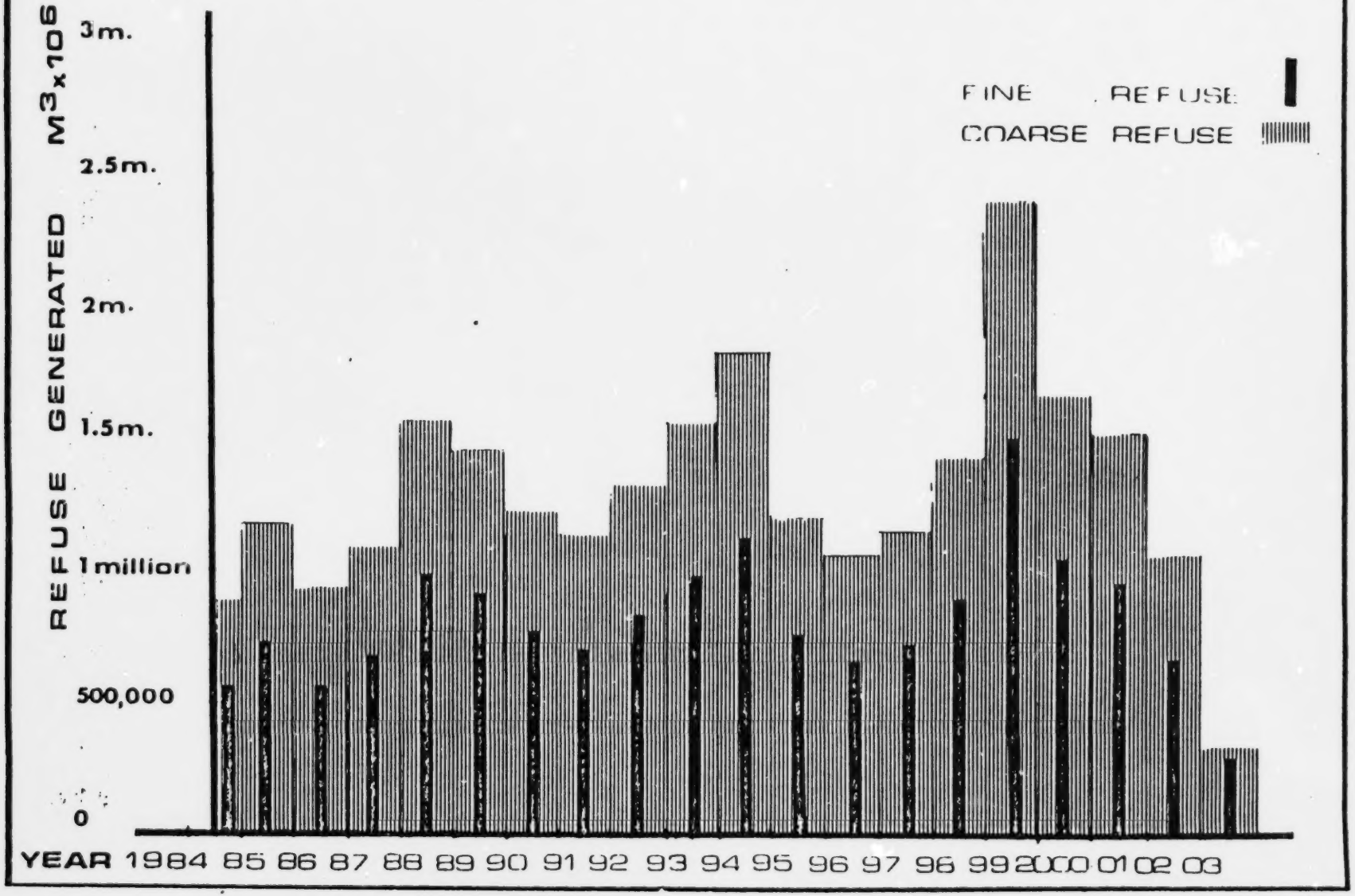
Lagoon A	250,000
Lagoon B	330,000
Lagoon C	1,192,000
Lagoon D	10,204,000
Total Fines	11,976,000 tonne

Coarse Refuse

North Extension 9/82	15,890,000
82-83 Solids Area 'B'	170,000
Extension	757,000
83-84	729,000
Total Coarse	17,546,000 tonne

FORECAST REFUSE GENERATION · 1984 To 2003

FIG. 2.1



According to Westar Mining Ltd.'s forecast mine plan to the year 2003, the daily production of coal rejects from the three selected streams will be maintained at a level slightly more than the requirements of the conversion plant. (Figure 2.1)

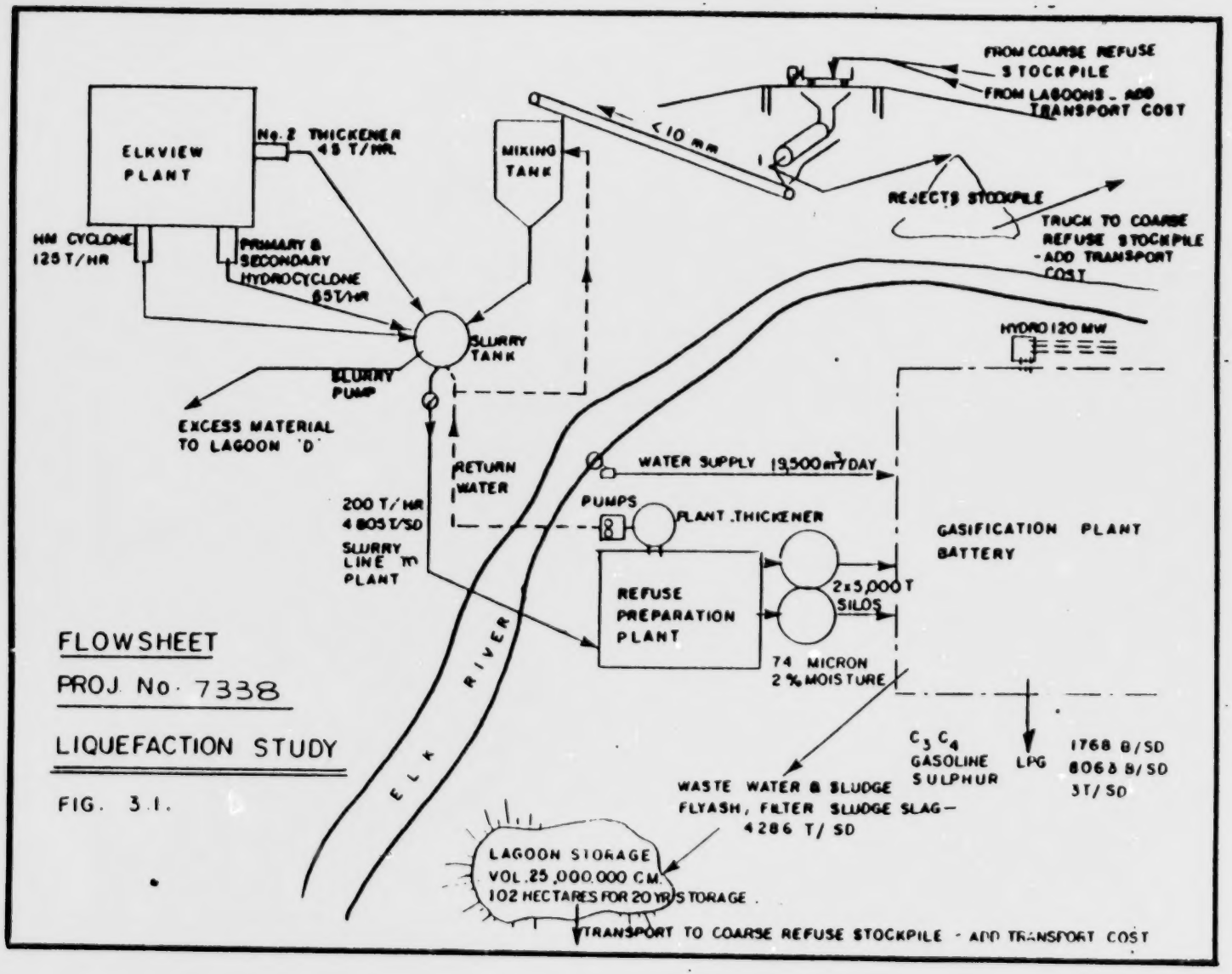
2.2 Feedstock Characteristics

The characteristics of the three feedstock streams of interest, i.e.:

- o Heavy Media Cyclone Refuse
- o Secondary Hydrocyclone Refuse
- o #2 Thickener Underflow

were identified in the Phase IA Report. The rejects from Barvoy Vessels are excluded as feedstock in the present study. As a result the composition is somewhat different and basically as follows:

STREAM	ANALYSIS OF COAL, AIR DRIED BASIS				
	Weight % in Composite	Moisture %	Ash %	Volatile Matter %	Fixed Carbon %
HM Cyclone Refuse	30	0.4	50.5	15.1	34.0
Secondary Hydrocyclone Refuse	30	0.6	37.1	17.6	44.7
No. 2 Thickener Underflow	40	1.4	28.5	18.2	51.9
Composite (by calculation)	-	0.8	37.7	17.1	44.4



FLOWSHEET

PROJ No. 7338

LIQUEFACTION STUDY

FIG. 3.1.

3.0 DESCRIPTION OF COAL TO GASOLINE COMPLEX

3.1 General

The main processing steps employed in the coal to gasoline complex are described in this section. Although several commercial processes are available for the various conversion steps, this report discusses the base case established in the Phase IA report. The main components of the processing sequence are as follows: feedstock preparation, coal gasification (Koppers-Totzek), methanol production (ICI) and gasoline production (Mobil).

3.2 Coal Preparation

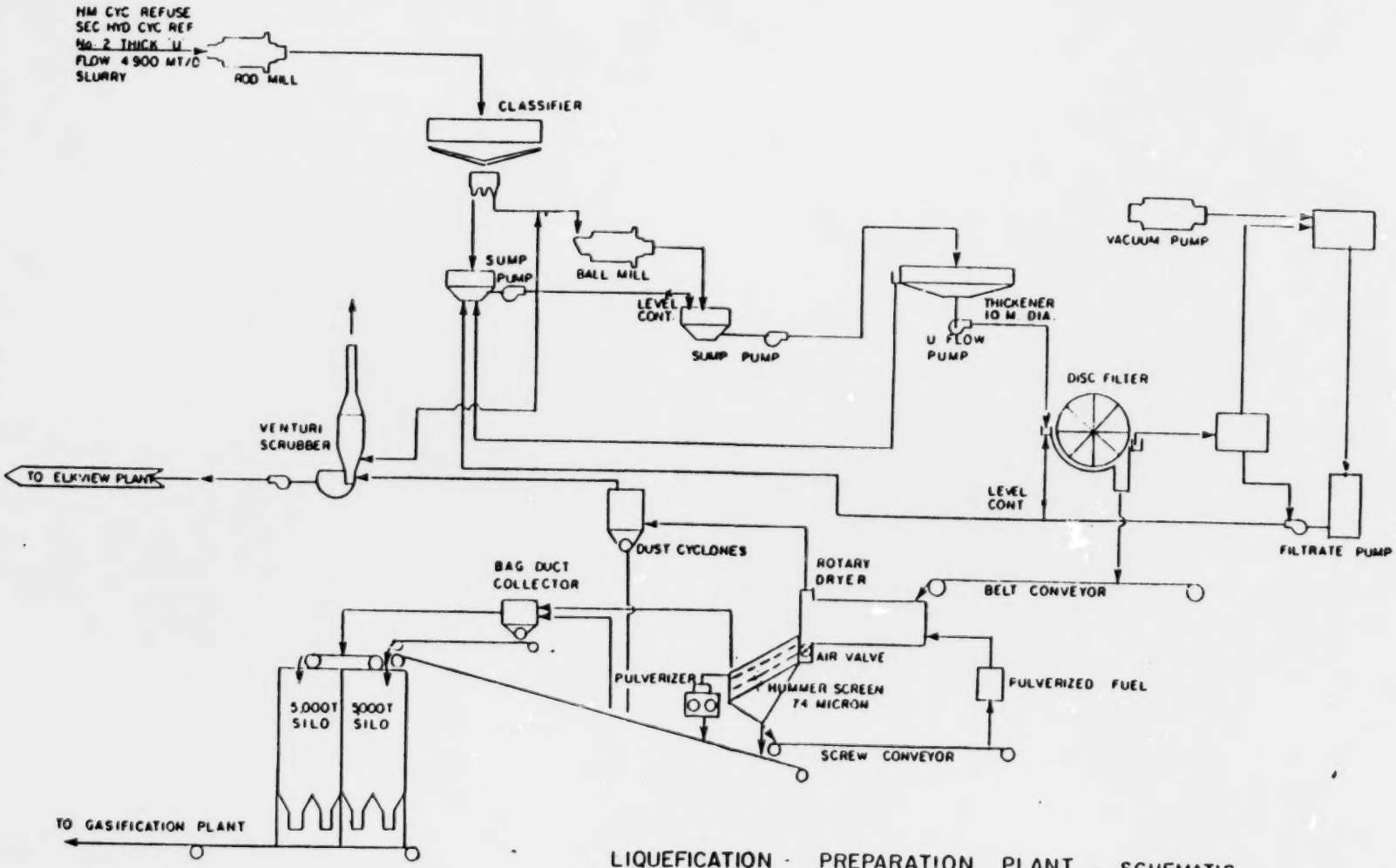
The coal reject streams selected for the process from the Elkview coal preparation plant are:

- Heavy Media Cyclone Refuse
- Secondary Hydrocyclone Refuse
- #2 Thickener Underflow

The three streams are received into a slurry tank at a site located near the Elkview Plant (see Figure 3.1). The combined material is mixed and pumped as a slurry across the Elk River to the Refuse Preparation plant (located next to the Gasification Plant) through a 25 cm dia. slurry line. The return water is pumped back to the slurry tank.

Slurrying is judged to be the most cost effective method of transport both from capital cost and operating cost considerations after considering alternatives such as conveying, trucking and utilizing a ropeway.

A bridge across the Elk River is specified to carry the slurry line, the return water line, and provide a vehicular link between the two locations.



LIQUEFICATION · PREPARATION PLANT - SCHEMATIC

FIG. 3.2.

Occasionally reject material is not available because the wash plant is not in operation over a period longer than for normal maintenance. When this occurs, material from the rejects stockpile is available to ensure uninterrupted operation of the Conversion Complex. In this case the rejects are trucked to a dump station from both the fines lagoons and the coarse coal stockpile. Rejects 9.5 mm and less are selected by a screening process and fed into a mixing tank. From this location the mixed material is slurried to the slurry tank and fed to the refuse preparation plant via the slurry line. The oversized material is trucked back to the stockpile. It is anticipated that roughly 30% of the stockpile refuse will be oversized.

The rejects feed stock is received at the refuse preparation plant in a mixed slurry form (see Figure 3.2), passes through a rod mill, and then is routed through a classifier to remove water. The rejects are next ground to a -0.075 mm size in a battery of ball mills. The ground material is routed via a thickener and disc filters where all of the fines are recovered and placed onto a conveyor which provides the feed to the rotary dryer. The dry material is pulverized and fed to two 5,000 tonne storage silos from where it is available as feed to the gasification plant. In the silos the gasifier plant feed stock is stored under an inert gas blanket.

As an alternative to the ball mills, the use of "micronizers" to prepare the reject material to -0.045 mm has been studied. This newer technology offers a $15 million saving in the overall project cost and is therefore used in the base case to provide the rejects preparation plant cost estimates.

A further alternative is designed to satisfy the feedstock needs for the Texaco gasifier option. In this case the final product is available in a slurry form at about 20% moisture content and is stored in tanks

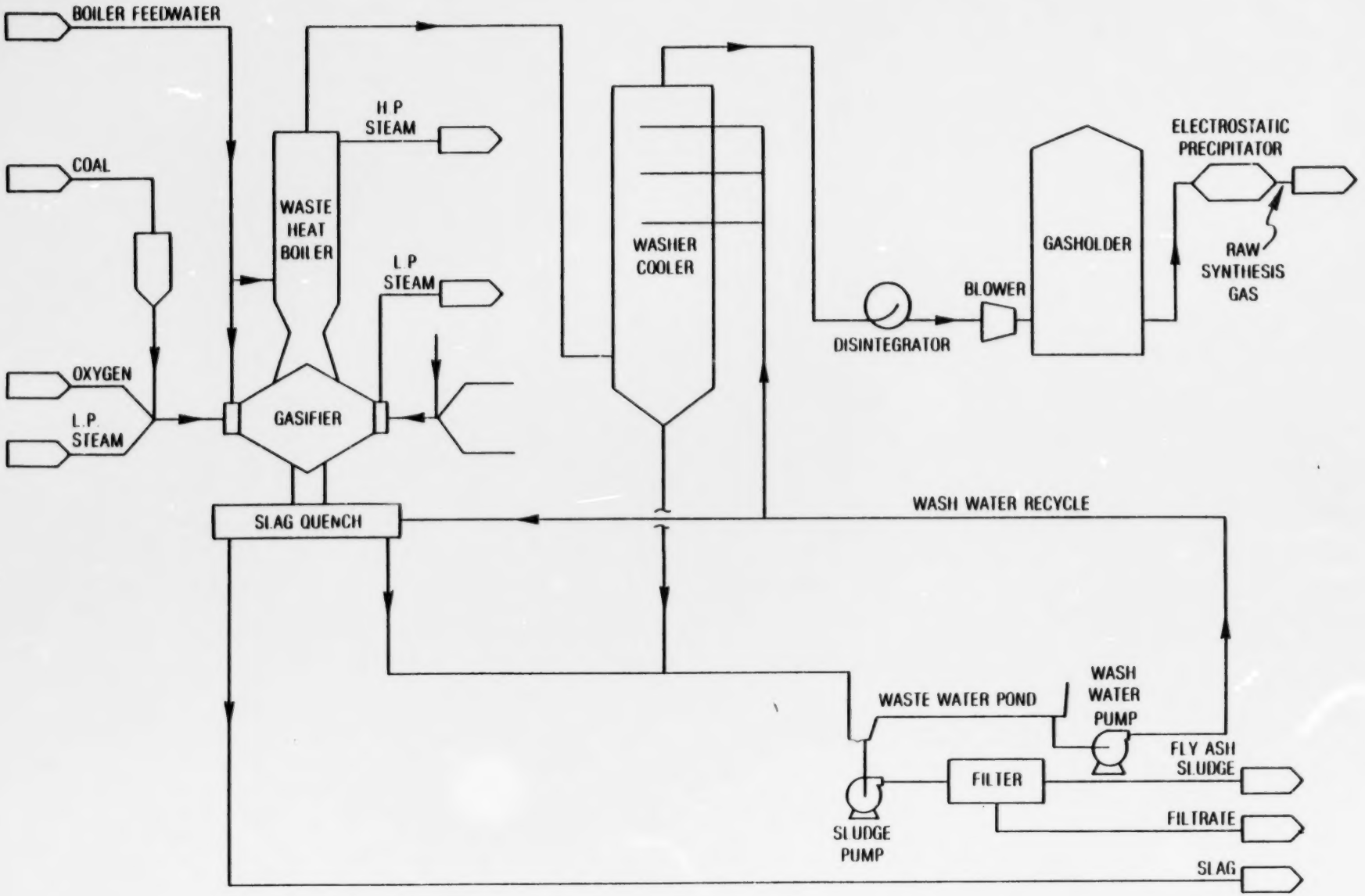


FIGURE 3.3

KOPPERS-TOTZEK GASIFICATION PROCESS

providing a total capacity of 1,000 tonnes. From these tanks the slurry is piped to the gasification plant. In the Texaco process the feed stock in this fluid state is acceptable.

3.3 Conversion Complex

A simplified flow plan of the coal conversion complex is shown in Figure 3.7. This complex is identical to the flow scheme presented in the Phase IA report.

3.3.1 Koppers-Totzek Gasification

The coal rejects feed to the complex moves by screw conveyors from the feed silos into a mixing zone where it is entrained in a pre-mixed stream of steam and oxygen and fed to the Koppers-Totzek (K-T) gasifiers (see Figure 3.3).

About 50% of the coal ash leaves the gasifier through the bottom in molten form and then is solidified by contact with water in a quench tank situated beneath the gasifier. The ash is removed from the quench tank by means of a scraper-conveyor assembly and is then conveyed to a slag storage bin which permits loading of the slag for disposal.

The remaining ash along with any ungasified carbon is entrained with the gas leaving the top of the gasifier. The gas stream is sprayed with water, if necessary, to solidify any slag droplets prior to entry into the waste heat boiler. High pressure saturated steam is generated in the waste heat boiler, and gas leaving the economizer section is directed to a washer-cooler. Here the gas is cooled by circulating water, typically to about 38°C, while about 90% of the entrained particulates are simultaneously removed. Gas then enters Theisen type irrigated disintegrators to reduce dust loading of the gas to about 4.6 mg/m^3. After passing through a mist eliminator, the gas is moderately boosted in pressure by a fan to overcome pressure losses. The K-T process consists of 7 gasifiers (6 on-line, 1 spare).

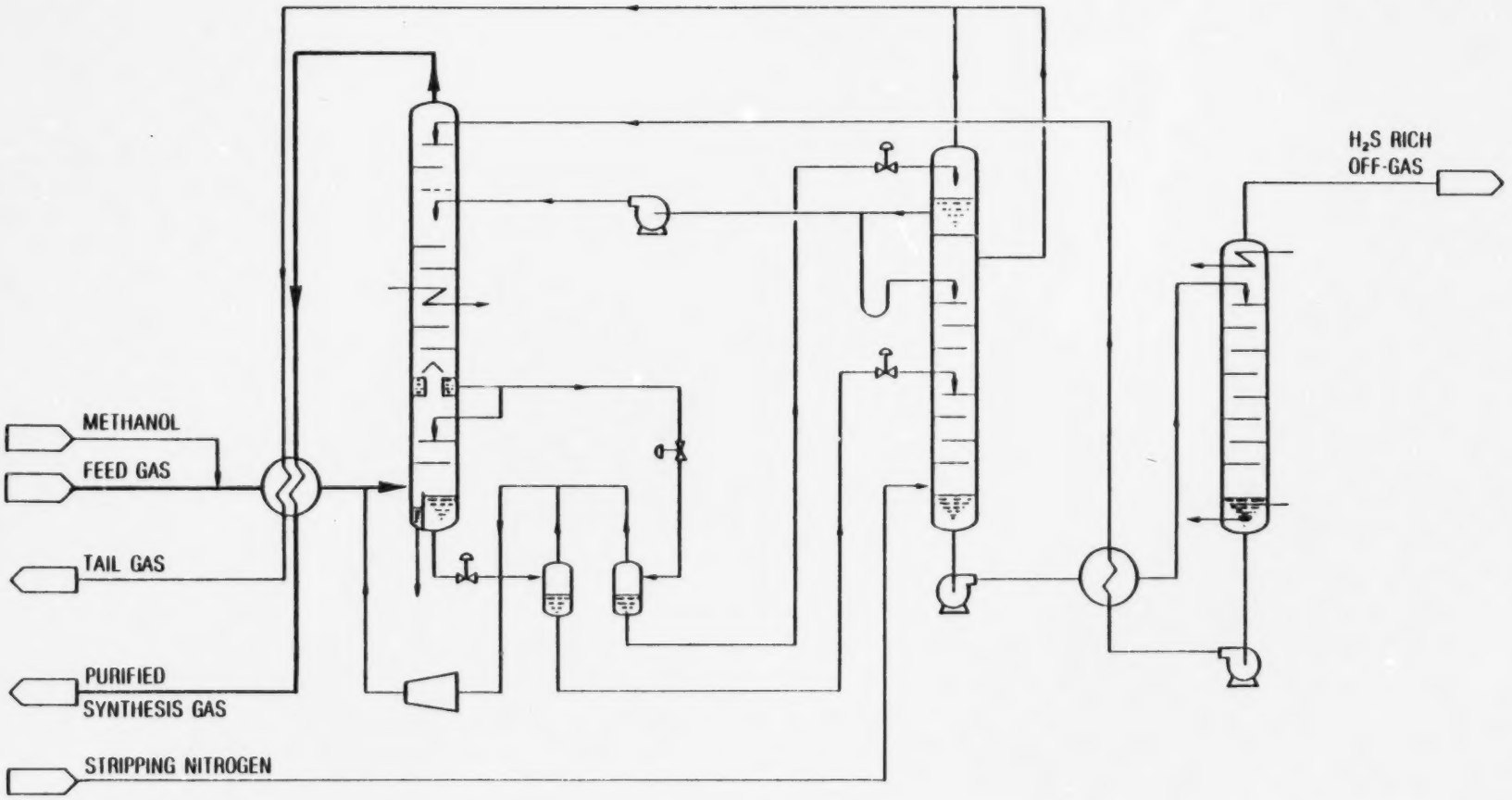


FIGURE 3.4
RECTISOL PROCESS

3.3.2 Oxygen Plant

Critical to the gasification process is the supply of oxygen. Two parallel trains are provided to supply high purity oxygen for use in the gasifiers, as well as nitrogen for the gasifiers, RECTISOL plant, MTG plant and product storage areas.

3.3.3 CO Shift

The CO shift conversion reactors produce hydrogen by the exothermic reaction of steam with carbon monoxide. In order to achieve the correct hydrogen to carbon monoxide ratio in the feed to the methanol plant, about 55% of the raw gas from the gasifiers is fed to the CO shift convertors and the remainder bypassed. A conventional CO shift unit contains both high temperature and low temperature shift convertors with extensive heat recovery. The shifted gas and the bypassed gas are combined and sent to the acid gas removal plant.

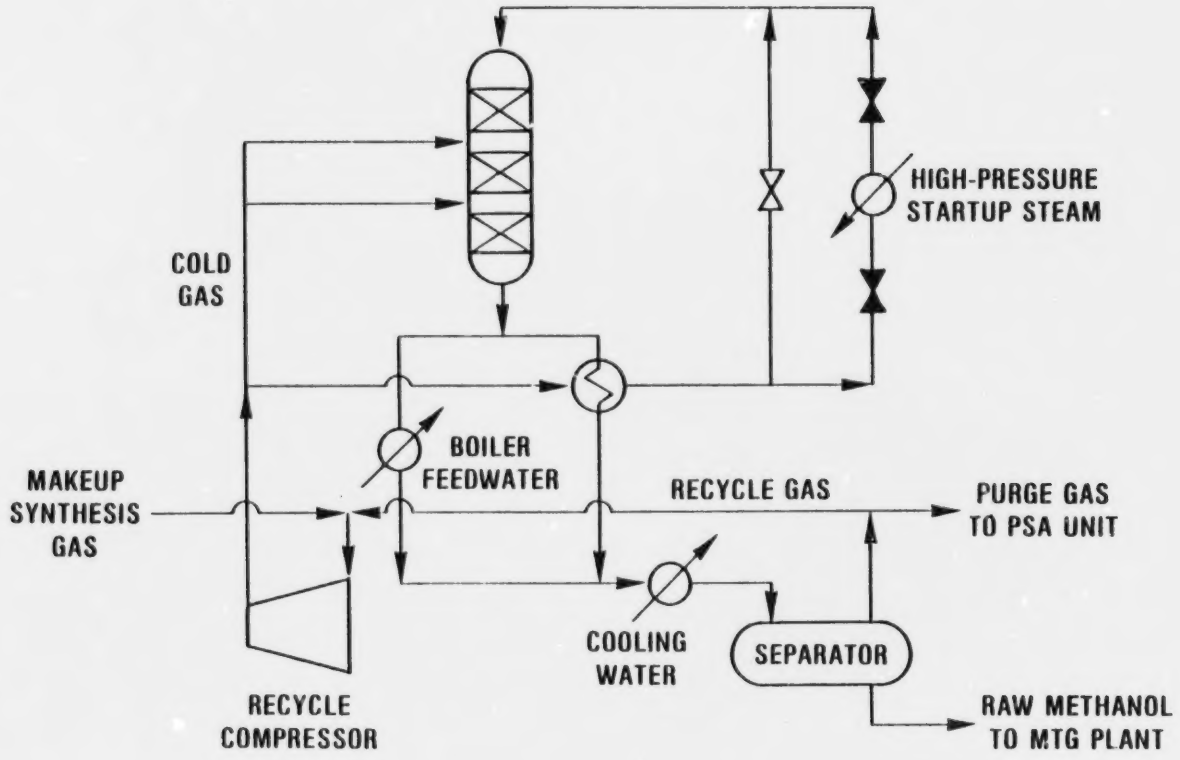
3.3.4 Acid Gas Removal

An acid gas removal plant is required to remove hydrogen sulphide, carbonyl sulphide and carbon dioxide from the synthesis gas feed to the methanol plant. The process used is RECTISOL using methanol as solvent (Figure 3.4). The sulphur compounds are removed by contact with cold CO_2 rich methanol. Carbon dioxide is removed to about 3% in the upper section of the column by contact with warm lean methanol. The purified gas is then passed to the methanol plant via the feed gas heat exchanger.

The RECTISOL plant contains a solvent regeneration system that produces two separate off-gas streams:

- Tail gas containing less than 5 ppm H_2S that is vented directly to atmosphere.

FIGURE 3.5
THE ICI
LOW PRESSURE METHANOL SYNTHESIS PROCESS



- H_2S rich off-gas that is sent to the Stretford Plant for the recovery of sulphur.

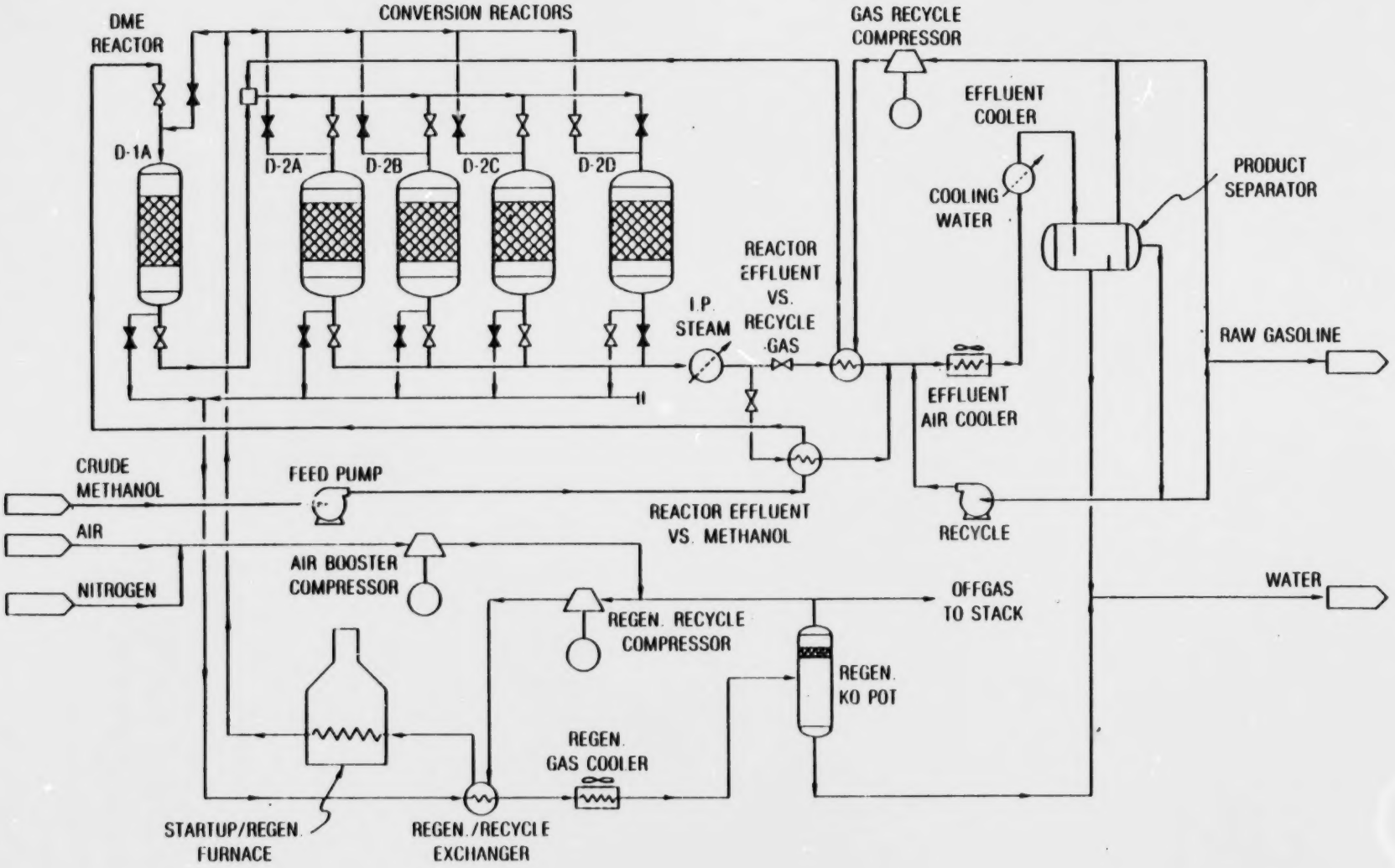
3.3.5 Sulphur Plant

The Stretford sulphur unit removes hydrogen sulphide from the concentrated RECTISOL plant acid gas effluent. The hydrogen sulphide (together with some carbon dioxide) is absorbed into the Stretford liquor, an aqueous solution of sodium carbonate, sodium meta vanadate and anthraquinone disulphonic acid (ADA). Clean gas from the absorber is vented directly to atmosphere. The absorbed hydrogen sulphide is converted to elemental sulphur by the reduction of vanadium.

3.3.6 Methanol Synthesis

Methanol is produced by the Imperial Chemical Industries (ICI) low pressure process (Figure 3.5). Carbon monoxide and hydrogen are combined over a copper based catalyst in an adiabatic reactor operating at 8400 kPa (120 psig). A recycle is required to achieve satisfactory overall conversion of the synthesis gas. A purge is taken from the recycle loop to prevent the accumulation of inerts (principally nitrogen, argon and methane). The purge gas is sent to a Pressure Swing Adsorption (PSA) unit for the recovery of hydrogen. Most of the hydrogen is recycled to the methanol plant and the remainder is sent to the heavy gasoline hydrotreaters in the Mobil MTG plant.

Raw methanol is condensed out of the converter product and then flashed to release dissolved gases. These gases are fed to the plant fuel gas system. The raw methanol produced in the synthesis loop contains about 5% water (derived from carbon dioxide in the synthesis gas) and about 2600 ppm higher alcohols and dimethyl ether. This methanol is a suitable feedstock to the Mobil MTG process and requires no further treatment.



SOURCE: MOBIL RESEARCH AND DEVELOPMENT CORPORATION

FIGURE 3.6
MOBIL MTG PROCESS, REACTION SECTION

3.3.7 Mobil Methanol to Gasoline (MTG) Process

The Mobil MTG process converts raw methanol into reasonably high octane gasoline and LPG. The process involves catalytic dehydration giving a stoichiometric yield of 44 wt% hydrocarbons and 56 wt% water (based on a pure methanol feed). The reaction is performed in two separate steps employing separate proprietary Mobil catalysts. The first stage converts methanol to dimethyl ether (DME) and water, liberating about 20% of the total heat of reaction. The second stage converts the DME to heavier hydrocarbons.

The reaction section is shown in Figure 3.6. Typically there are two conversion reactors on-line with a third reactor being regenerated.

The reactor product is cooled and separated into water, gas and liquid hydrocarbons. The water is sent to a treatment plant and some of the gas is recycled. The remaining gas and the liquid hydrocarbons are sent to product fractionation. The fractionation section produces; fuel gas, propane (C_3), butane (C_4), light gasoline and raw heavy gasoline. The raw heavy gasoline contains durene (1,2,4,5-tetramethyl benzene) a high freezing point impurity. The durene concentration is reduced by mild hydrotreating in the heavy gasoline treating (HGT) unit. The hydrogen feed to the HGT unit is a portion of the purified methanol loop purge gas. The finished gasoline products are sent to product blending. The MTG gasoline is a high octane material as shown in Table 3.1.

Typical compositions of the LPG byproducts are given in Table 3.2.

TABLE 3.1

Estimated Properties of Blended Gasoline

Gravity, Kg/m^3	722 (64° API)
Octane Numbers	
Research, Clear	93
Motor, Clear	83
(R+M)/2, Clear	88
Reid Vapor Pressure, kPa	69 (10 psi)
Distillation, °C	
IBP	31
10%	47
30%	69
50%	100
70%	138
90%	156
EP	204
Composition, vol. %	
Paraffins	52
Olefins	16
Naphthenes	5
Aromatics	27
Durene Content, wt%	2
Molecular weight	94

TABLE 3.2

Typical Propane/Butane LPG Compositions

Component	Propane Vol. %	Mixed Butanes Vol. %
Ethane	4.0	-
Propylene	3.7	-
Propane	90.5	0.1
Isobutane	1.7	74.2
Butenes	0.1	8.1
n-Butane	-	17.4
Pentanes	-	0.2

3.3.8 Utilities Systems

The plant requires four major utility systems. These are:

- Cooling water system. A closed circuit cross-flow induced draft cooling tower is employed.

- Water treatment. This unit includes waste water treatment, boiler feed and process water production sections. There is a substantial import of raw water (13-15 kl/min) to compensate for the evaporative loss from the cooling tower and the consumption of process steam (particularly in the gasifier and CO shift reactors). Two qualities of water are produced, boiler feedwater and process water. The process water is used as make up to the gasifier wash water system and the cooling tower circuit.

- A steam boiler/superheater. This gas fired unit is oversized so that it can also serve as a start-up unit.

- Refrigeration, which is required for the RECTISOL plant.

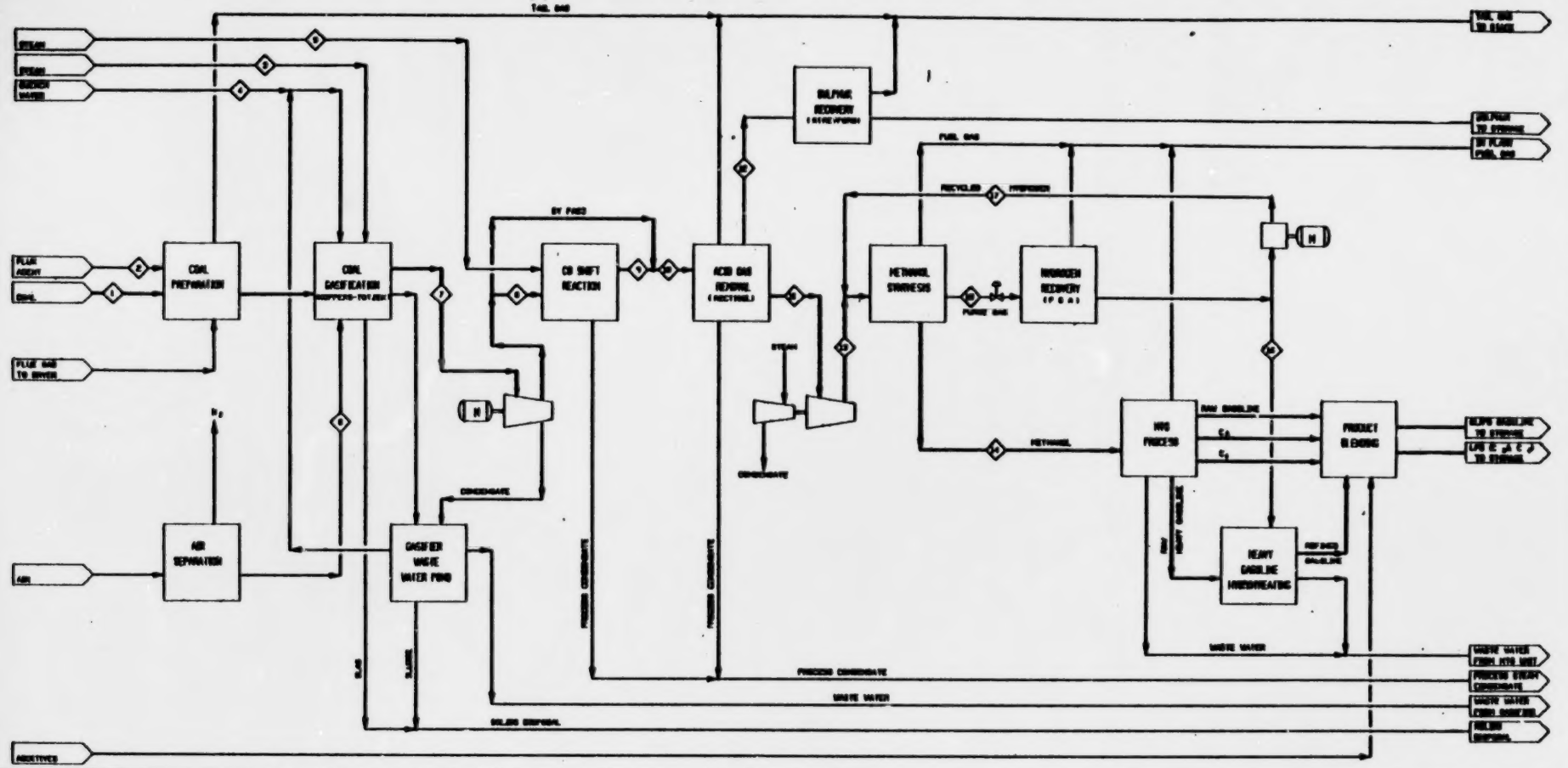
3.3.9 General Offsites

The complex contains the usual general offsites facilities such as:

- Relief and blowdown systems
- Flares
- Intermediate and product storage
- Inter-connecting piping
- Electrical distribution
- Fire protection
- Sewers
- Roads
- Buildings
- Workshops

Figure 3.7

Simplified Flow Plan and Material Balance



STREAM NO.	1	2	3	4	5	6
UNIT	t/d	t/d	t/d	t/d	t/d	t/d
C	2635.5					
H	143.5					
N	31.2					
S	2.4					
O	241.9					
Cl	0.35					
Ash	1,945.45					
H_2O	261.4		563.6	1,720.5	3,399.0	
Sulphur		4.54				
H_2 (99% purity)						2,829.4
TOTAL	5,061.70	4.54	563.6	1,720.5	3,399.0	2,829.4

STREAM NO.	7		8		9		10		11		12		13		14		15		16		17	
UNIT	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d
CO_2	799.2	18.1	434.8	9.9	4,360.4	99.1	4,724.8	107.4	314.6	7.1	159.2	3.62	314.6	7.1	1.47	0.04	21.4	0.49				
CO	4,594.1	164.1	2,499.2	89.3			2,094.9	74.8	2,080.4	74.3			2,080.4	74.3			60.9	2.5				
H_2	157.9	79.0	85.9	43.0	264.4	132.2	336.4	168.1	335.5	167.8			335.5	167.8	0.04	0.02	55.7	27.9	1.19	0.60	48.9	24.45
N_2	49.4	1.8	26.9	0.96	26.9	0.96	49.5	1.8	49.5	1.8	4.51	0.15	49.5	1.8			49.4	1.8				
Ar	35.8	0.9	19.5	0.48	19.5	0.49	35.8	0.9	35.8	0.9			35.8	0.9			35.8	0.9				
H_2S	2.7	0.08	1.4	0.04	1.4	0.04	2.62	0.08			2.62	0.08										
CH_4	0.42	0.03	0.23	0.01	0.23	0.01	0.42	0.026	0.42	0.026			0.42	0.026			0.42	0.026				
HCl	0.98	0.03	0.53	0.01	0.53	0.01	0.44	0.01														
H_2O	287.1	15.9	2.3	0.12	5.6	0.20	3.7	0.21							119.0	6.66	0.07	0.004				
CH_3OH															2,504.0	78.3	5.92	0.19				
TOTAL	5,927.6	280.0	3,070.7	143.7	4,677.1	233.0	7248.6	351.3	2,816.1	251.9	166.2	3.85	2,816.1	251.9	2,625.5	85.0	237.6	33.7	1.19	0.60	48.9	24.45

In addition to the storage for the gasifier raw gas and the MTG intermediate products, additional storage is provided as follows:

- o Liquid oxygen: 1,500 t (12 hours)
- o Gaseous nitrogen: 75,000 m^3 (4 hours)
- o Crude methanol: 24,000 m^3 (1 week)
- o MTG products: 22,000 m^3 (2 weeks)
- o Miscellaneous: 4,000 m^3

3.4 Material Balances

The material balance for the base process scheme is shown in Table 3.3. The rates summarized in this table are one-half the rates shown in the Phase IA report. The numbered streams from Table 3.3 are identified on Figure 3.7.

3.5 Utility/Energy Balances

Tables 3.4 and 3.5, summarize the utility and overall energy balances, respectively. Again, the utility balance represents about one-half of the Phase IA report utility balance. The energy balance and reported overall energy efficiency are calculated specifically for this report.

TABLE 3.3

Coal to Gasoline Complex Material Balance

*all days are stream days unless otherwise noted

STREAM NO.	1	2	3	4	5	6
	t/d*	t/d	t/d	t/d	t/d	t/d
C	2435.5					
H	143.5					
N	31.2					
S	2.4					
O	241.9					
Cl	0.35					
ASH	1,945.45					
H_2O	261.4		563.6	1,720.5	3,399.0	
FLUX		4.54				2,829.4
O_2 (99% purity)						
TOTAL	5,061.70	4.54	563.6	1,720.5	3,399.0	2,829.4
TEMP. (°C)	25	25	135	25	285	25
PRESS (kPa)	86.2	86.2	300	300	7,000	89.1

TABLE 3.3 (Cont'd)

STREAM NO.	7		8		9		10		11	
UNIT	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d
CO_2	799.2	18.1	434.8	9.9	4,360.4	99.1	4,724.8	107.4	314.6	7.1
CO	4,594.1	164.1	2,499.2	89.3			2,094.9	74.8	2,080.4	74.3
H_2	157.9	79.0	85.9	43.0	264.4	132.2	336.4	168.1	335.5	167.8
N_2	49.4	1.8	26.9	0.96	26.9	0.96	49.5	1.8	49.5	1.8
Ar	35.8	0.9	19.5	0.48	19.5	0.49	35.8	0.9	35.8	0.9
H_2S	2.7	0.08	1.4	0.04	1.4	0.04	2.62	0.08		
CH_4	0.42	0.03	0.23	0.01	0.23	0.01	0.42	0.026	0.42	0.026
HCl	0.98	0.03	0.53	0.01	0.53	0.01	0.44	0.01		
H_2O	287.1	15.9	2.3	0.12	3.6	0.20	3.7	0.21		
CH_3OH										
TOTAL	5,927.6	280.0	3,070.7	143.7	4,677.1	233.0	7248.6	353.3	2,816.1	251.9
TEMP. (°C)	35		179		35		35		25	
PRESS (kPa)	101.7		6,900		6,800		6,800		6,650	

TABLE 3.3 (Cont'd)

12		13		14		15		16		17	
t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d	t/d	10^3kmol/d
159.2	3.62	314.6	7.1	1.67	0.04	21.4	0.49				
		2,080.4	74.3			68.9	2.5				
		335.5	167.8	0.04	0.02	55.7	27.9	1.19	0.60	48.9	24.45
4.31	0.15	49.5	1.8			49.4	1.8				
		35.8	0.9			35.8	0.9				
2.62	0.08										
		0.42	0.026			0.42	0.026				
				119.8	6.66	0.07	0.004				
				2,504.0	78.3	5.92	0.19				
166.2	3.85	2,816.1	251.9	2,625.5	85.0	237.6	33.7	1.19	0.60	48.9	24.45
25		50		35		35		35		116	
200		8,274		86.3		7,763		2,515		9,274	

TABLE 3.4
Coal to Gasoline Complex Utility Balance

	Coal Preparation(1)	Air Separation	Gasification	Raw Gas Compression	CO Shift	Acid Gas Removal(2)	Methanol Plant(3)	MTG Plant	Utilities and Offsites (4)	Total
STEAM, 100 bar SATURATED, t/d										
Produced	0	0	4,690	0	0	0	0	0	0	4,690
Consumed	0	0	0	0	0	0	0	0	4,690	4,690
STEAM, 100 bar SUPERHEATED, t/d										
Produced	0	0	0	0	0	0	0	0	2,460	2,460
Consumed	0	0	0	0	0	0	420	1,231	809	809
STEAM, 70 bar SATURATED, t/d										
Produced	0	0	0	0	1,262	0	0	0	2,578	3,840
Consumed	0	0	0	0	3,399	60	0	381	0	3,840
STEAM, 3 bar SATURATED, t/d										
Produced	0	0	1,927	0	1,892	0	0	84	0	3,903
Consumed	0	0	563	0	0	268	0	0	3,072	3,903
ELECTRIC POWER, MW (5)										
Produced	0	0	0	0	0	0	0	0	9.3	9.3
Consumed	8	34	7	60	0	1.9	2.2	4.4	14.8	132.3
COOLING WATER, $10^3 m^3/d$ (6)										
Produced	0	0	0	0	0	0	0	0	858	858
Consumed	0	83	5	95	0	12	215	221	227	858
PROCESS WASTE WATER, m^3/d										
Produced	0	0	670	0	1,793	0	0	1,640	0	4,103
Consumed	0	0	0	0	0	0	0	0	4,103	4,103
BLOWDOWN STREAMS, m^3/d										
Produced	0	0	135	0	63	0	0	57	1,726	1,981
Consumed	0	0	0	0	0	0	0	0	1,981	1,981
CONDENSATE, m^3/d										
Produced	0	0	0	0	0	328	420	185	3,512	6,520
Consumed	0	0	0	0	0	0	0	0	6,520	6,520
PROCESS WATER, m^3/d										
Produced	0	0	0	0	0	0	0	0	20,629	20,629
Consumed	0	0	1,720 (P)	0	0	0	0	0	18,809	20,629
BOILER FEED WATER, m^3/d										
Produced	0	0	0	0	0	0	0	0	10,798	10,794
Consumed	0	0	6,753	0	3,217	0	0	387	437	10,794
FUEL GAS, GJ/d (7)										
Produced	0	0	0	0	0	0	1,706	1,459	0	3,165
Consumed	0	0	0	0	0	0	0	0	3,165	3,165

NOTES: 1) Energy for coal drying will be recovered from either surplus LP steam or boiler flue gas
2) Utilities for both RECTISOL and Stretford plants included
3) Utilities for both Methanol synthesis loop and hydrogen recovery and recycle included
4) In the utilities plant 2777 t/d of surplus LP steam is condensed against cooling water
5) Imbalance i.e., 123 MW is imported
6) Cooling water temperature rise of 10°C
7) Fuel gas utilization at 85% thermal efficiency is assumed
8) Process consumption of steam and water denoted (P)

TABLE 3.5

Koppers-Totzek Energy Balance

		Terajoules/CD (HHV)	
In:			
	Coal Feed	91.18	
	Power	30.12*	
	Total	121.30	
Out:			
	MTG Gasoline	38.69	
	LPG's	6.90	
	Sulphur	0.02	
	Fuel & Losses	75.69	(By diff.)
	Total	121.30	

$$\text{Overall Energy Efficiency} = \frac{\text{MTG Gasoline + LPG + Sulphur}}{\text{Coal Feed + Power}} \times 100$$

$$= \frac{38.69 + 6.90 + 0.02}{91.18 + 30.12} \times 100$$

$$= \frac{45.61}{121.30} \times 100$$

$$= 38\%$$

* 1 KWH = 3.6 mj generated @ 34% efficiency = 10.55 mj

TABLE 4.1

Phase IA Gasifier Comparison at Half Scale

	Texaco	Koppers-Totzek
Feed:		
Coal Feed t/sd	4,800	4,800
Oxygen Feed t/sd	2,970	2,830
Water Feed t/sd	3,720	1,580
Products:		
Methanol Produced t/sd	2,840	2,625
Gasoline Produced m^3/sd	1,470	1,280
LPG Produced m^3/sd	300	280
Total Saleable Production m^3/sd	1,770	1,560
Ash + Slag Produced t/sd	4,150	3,676

4.0 COMPARISON OF GASIFICATION PROCESSES

4.1 General

As noted in Section 1.0, one of the tasks is to update the gasification process selection results. Exxon Research and Engineering Co. (ERE) provided a re-evaluation and update of the comparison between the Koppers-Totzek and Texaco technologies given the progress of the Texaco gasification process development.

4.2 Process Comparison

ERE's review of the Phase IA Texaco and K-T gasification technology comparison draws different conclusions. ERE recommend that future studies be based on the Texaco process which has been successfully commercialized since the original Phase IA report evaluation in 1982.

The overall conversion of coal to gasoline based on these two gasifiers is summarized in Table 4.1. The Texaco gasifier produces a higher liquid hydrocarbon yield per tonne of coal with a slightly higher consumption of oxygen and process water.

The Phase IA report developed a comparison of the Texaco and K-T gasification processes based on:

- process factors (e.g. temperature, pressure)
- technical risks (e.g. feeding, slag removal)
- commercial factors (proven experience)

The results of this comparison are shown in Tables 4.2-4.4. The numerical assessment of these factors weighed in favor of the K-T gasifier. The principal factors favoring K-T were lower technical risk, higher service factor, and proven commercial operation vs. Texaco's limited demonstration plant experience three years ago at the time of the Phase IA analysis.

TABLE 4.2

Gasifier Comparison, Process Factors*

		Texaco		Koppers-Totzek	
Process Factors	Weight	Value/Comments	Rate	Value/Comments	Rate
Temperature (°C)	.05	1460	8	1500	8
Pressure (kPa/psig)	.20	5860/850	7 **	Atmospheric	10
Oxygen Consumption (kg/kg Coal)	.20	0.68	7	0.64	7
Steam Consumption (kg/kg Coal)	.20	0.415***	5(10)	0.19	9
Carbon Conversion(%)	.20	97	10	90	8
Coal Preparation	.10	Coal pulverized Slurry preparation Flux additive	5(8)	Coal pulverized	8
Feedstock Preheating (°C)	.05	200	8(10)	Not required	10
Subtotal	1.00	W X R	7(8.4)	W X R	8.5

Notes:

* Numerical values in parenthesis represent ERE's current assessment, while base values were estimated for the Phase IA report.

** The higher operating pressure of the Texaco Unit is an advantage in relation to downstream operations. The lower rating here reflects the potential operating problems that the higher pressure will cause in the gasifier and immediate ancillary equipment.

*** Consumed as water in the Texaco process.

TABLE 4.3

Gasifier Comparison, Technical Risks*

		Texaco		Koppers-Totzek	
Technical Risks	Weight	Comments	Rate	Comments	Rate
Feed System	.20	Troublesome slurry pumps (not proven on a commercial scale)	6(10)	Conventional screw conveyor proven on a commercial scale	10
Reactor - Construction Matr'l	.25	Refractory lined Refractory life-time 6-9 months	5(8)	Refractory lined Refractory life-time 12-14 months	8
- Burners	.25	1 Burner/Gasifier - Abrasion - Must shutdown unit when problems arise	5(10)	4 Burners/Gasifier - Small burners - Unit can operate with only 2 burners	10
Slag Removal	.15	Problem encountered in ash removal	6	No problem in slag removal but high amount of fly ash to be removed after gasification	8
Heat Recovery	.15	No operating experience with waste heat boilers	5(10)	Waste heat boilers available	10
Subtotal	1.00	W X R	5.4(9.0)	W X R	9.2

* Numerical values in parenthesis represent ERE's current assessment, while base values were estimated for the Phase IA report.

TABLE 4.4
Gasifier Comparison, Commercial Factors*

		Texaco		Koppers-Totzek	
Service Factor	Weight	Comments	Rate	Comments	Rate
Service Factor (%)	1.00	80(95)	7(9.5)	95	9.5
Subtotal	1.00	W X R	7(9.5)	W X R	9.5
Environmental Considerations					
Discharges to Air	.33	Possible contamination from flash drums and stripping components in scrub water	8.5	Possible contamination from stripping components in scrub water (COS, HCN, SO_2 & NO)	8.5
Discharges to Water	.33	Water treated & recycled	8.5(10)	Water treated & recycled	8.5
Discharges to Land	.33	No hazard	10	No hazard	10
Subtotal	1.00	W X R	9(9.5)	W X R	9
Proven Experience					
Unit Scale	.75	15 t/d Pilot Plant 190 t/d Demo Plant developed problems. 3 units under construction **(900 t/d)	4(8)	Commercial unit 40 installed 600-1200 t/d	10
Expected Date of Commercialization	.25	1986**	8(10)	Commercial	10
Subtotal	1.00	WXR	5(8.5)	WXR	10

* Numerical values in parenthesis represent ERE's current assessment, while base values were estimated for the Phase IA report.

** Three commercial scale units now in successful operation.

When the numerical values for the factors are updated based on the current status of both processes, they show no advantage for K-T. (Updated values are shown in parentheses in Tables 4.2-4.4). The major change, since the Phase IA comparison, is the successful commercialization of the Texaco process. There are now commercial scale Texaco gasifiers operating at Tennessee Eastman and Cool Water in the U.S., and at Ube in Japan.

Updated process factors include:

- Steam consumption - Texaco does not require any steam feed but produces a great deal of high pressure steam.
- Coal preparation - Texaco can use wet grinding, a cheaper, simpler system than dry grinding with inert storage.
- Feedstock preheating - Texaco uses a simple step.

These adjustments make the process comparison breakeven.

Updated technical risk factors include:

- Feed system - Texaco slurry pumps are now proven on commercial scale. Slurry pumps require higher maintenance than K-T screw feeders. However, K-T feed nozzles require complicated balancing during operation.
- Reactor construction material - Texaco refractory is also expected to last 12 months.

- o Reactor burners - Abrasion problems are the same for both gasifiers. K-T reactors must operate symmetrically so if one burner fails the opposite burner must be turned off. Since repairs cannot be made while the remaining burners are in operation, the reactor can only run at one-half throughput, and must eventually be completely shut down to make burner repairs.

- o Heat recovery system - Texaco waste heat boilers have been operated at Oberhausen, Germany (large pilot scale) and at Cool Water in the U.S.

These adjustments make the gasifier technical risk comparison breakeven.

Updated commercial factors include:

- o Service factor - With either K-T or Texaco gasifiers, high service factor is obtained by providing a spare gasifier.

- o Environmental considerations/discharges to water - Texaco can incinerate any hazardous byproducts in wastewater streams by using the wastewater to slurry the coal.

- o Proven experience - First commercial Texaco unit started up in 1983 and three commercial scale units are currently operating.

These adjustments make the commercial experience comparison much closer. Most of the installed K-T plants have been shut down.

TABLE 4.5

Gasifier Comparison, Overall Evaluation*

	Weight	Texaco	Koppers-Totzek
Process Factors	1.0	7.0 (8.4)	8.5
Technical Risks	3.5	5.4 (9.0)	9.2
Commercial Factors			
- Service Factor	3.5	7.0 (9.5)	9.5
- Environmental	1.0	9.0 (9.5)	9.0
- Proven Experience	1.0	5.0 (8.5)	10.0
TOTAL	10	64 (91)	93

* Numerical values in parenthesis represent ERE's current assessment, while base values were estimated for the Phase IA report.

The overall gasifier comparison using these factors is shown in Table 4.5. The difference in the total weighted factor (91 for Texaco versus 93 for K-T) is well within the accuracy of the comparisons made. Therefore, this updated process comparison shows no advantage for K-T.

In addition to ERE's analysis, two U.S. government agencies, the Department of Energy (DOE) and the Tennessee Valley Authority (TVA), have sponsored independent analyses comparing Texaco and K-T as syngas producers. A 1980 study by C.F. Braun and Co. for TVA showed a 20% operating cost savings and a 5% investment savings for Texaco over K-T. The executive summary of that report is included in Section 13.1. Likewise, a 1984 study by Oak Ridge National Lab for DOE showed a 13% product cost savings and 10% investment savings for Texaco over K-T on bituminous coal. The executive summary of that report is found in Section 13.2.

The Electric Power Research Institute (EPRI) has also sponsored extensive coal gasification studies. A 1983 report for EPRI by Synthetic Fuels Associates[1] points out K-T's disadvantages due to operating at atmospheric pressure.

The studies by Braun and Oak Ridge National Lab show an economic advantage for Texaco over K-T in gasifying bituminous coals. Also, the update of the Phase IA process comparison shows Texaco has caught up with K-T's commercial experience. Therefore, future studies of gasification of high ash coal rejects should be based on Texaco gasifiers.

---

(1) "Raw fuel gas generation at low pressure poses a number of economic and operational problems. First, it is necessary to recover sensible heat from the high temperature raw gas prior to removing the fly ash and sulphur-bearing compounds. The heat transfer equipment necessary to cool low-pressure gas will be larger and more costly than that required for high-pressure gas. Second, particulate removal at low pressure is substantially less efficient than at high pressure. Finally, the gas pressure is generally elevated prior to the sulphur removal step. Compressing the K-T gas to, say, 450 psig adds significantly to the plant cost and auxiliary power requirement."

5.0 INFRASTRUCTURE & SUPPORT FACILITIES

5.1 General

In this stage of the Study a preliminary identification of the support facilities required specific to Site E are made. In the process of selecting this location for the Conversion Complex and the Rejects preparation plant, three other sites are also considered, namely, the Grave Prairie Site, Michel Site and the Olson/Hosmer Site (see Figure 5.1). Site E has the following characteristics that make it attractive and the choice for the plant location.

- o Close proximity to the Elkview Wash Plant rejects lagoons and stock-piles.
- o Adequate land availability for all plant and utility installations.
- o Adequate land availability for refuse disposal and for the solid and liquid wastes from the Conversion Complex.
- o Remote from highway and population centres, yet has excellent access.
- o Rail loop.
- o Proximity to a suitable source of water.
- o Relatively flat piece of terrain offering low development cost exposure.

5.2 Other Support Facilities

In addition to the general offsites associated with the Conversion Complex (see Section 3.3.9) the following support facilities are also necessary and included:

- o River intake, pumping system and water line
- o Ash disposal lagoons
- o Interplant conveyors
- o Railroad sidings
- o Reject slurry mixing tanks, slurry pumps and slurry pipe line
- o Rejects dump station

6.0 ENVIRONMENTAL CONSIDERATIONS

6.1 General

A preliminary contact with the B.C. Pollution Control branch was made to discuss the overall requirements and the need to comply with the B.C. Pollution Control regulations. Prior to the detailed design stage of the scheme, detailed drawings and complete chemical, physical and geological characteristics of the effluents and emissions have to be provided to the Pollution Control Inspectorate together with formal disposal applications.

6.2 Potential Impacts

6.2.1 Solid Effluents

These include:

- Coarse refuse from the coal preparation plant.
- Slag from the gasifier.
- Filter sludge from the gasifier waste water pond.
- Spent catalyst from the CO shift, methanol and MTG plants (dehydration and HGT units).
- Sulphur, if there is no market for sales.
- Sludge from the water treatment plant.

6.2.2 Liquid Effluents

All liquid effluents are routed to the water treatment plant. Under normal conditions the only liquid plant effluent is concentrated waste suitable for disposal to evaporation ponds and holding lagoons.

6.2.3 Gaseous Emissions

These include:

- Fugitive dust emissions from the coal preparation plant.
- Exhaust air from the coal driers which may contain particulates.
- Residual H_2S emission.
- MTG catalyst regeneration off-gas.
- Steam boiler flue gas.
- Cooling tower plume.
- Process and storage vent losses.
- Emergency reliefs to flare.

6.2.4 Miscellaneous

The complex incorporates some large items of mechanical equipment which require suitable noise control measures. The rejects preparation plant has suitable dust collection and disposal systems. The raw water make-up demand is about 19.5 million litres per day. This water is taken from the Elk River. The Water Rights Branch indicates that the volume is available.

7.0 CAPITAL COST ESTIMATE

7.1 General

The third phase of this study (see Section 1.3) is to re-evaluate the previous Phase IA cost estimate at one-half the plant size and on a current economic basis. ERE was hired to evaluate the Phase IA results and to advise on the costing methodology for a half-size plant. The review includes an examination of the overall scope and basis, an evaluation of the estimating methodology, development of overall and individual process block check estimates and a comparison of economic bases with ERE data. Proration factors to facilitate scaling the original 9,600 t/sd estimate down to a half size plant are also developed.

7.2 ERE Cost Estimate Analysis of Phase IA Report

7.2.1 Summary of Observations

Based on comparisons with studies having similar processing schemes, it appears that the most probable cost estimate for the complete 9,600 t/sd project is in the range of two to three times the Phase IA estimate. Major reasons for this increase include:

- The estimating methodology employed in Phase IA tended to underestimate the cost of the basic facilities.
- The estimate did not cover the full scope of facilities and services required.
- The project contingency level overstated the stage of project development.
- Owner's costs were not complete, and the infrastructure investment was excluded.

7.2.2 Approach to Cost Review

A review of the Phase IA study report reveals very limited information on the project investment. A meeting was held with PetroCanada to review the project on a technical and cost basis. Although Petro-Canada is no longer involved in the project, they were willing to set aside one day to provide background. While they were able to qualitatively discuss the estimate methodology and content, few additional quantitative details were available.

The lack of basis and estimating details limited the review to three major aspects: (1) developing an understanding of the estimating methodology employed by Petro-Canada, (2) comparing the overall investment with data for comparable projects and checking a few of the major unit costs against in-house data, and (3) reviewing the scope and content of the estimate for completeness. The specific conclusions in each of these areas are discussed below.

7.2.3 Estimate Methodology

Petro-Canada obtained estimates from licensors, vendors and contractors for 75% of the total project investment. The remainder was primarily based on in-house estimates. The outside cost data was generally based on turn-key battery limit units for a specific location such as U.S. Gulf Coast, U.S. Mid-West or West Germany. The Petro-Canada location factors used to transfer these costs to a British Columbia site appear low when compared with ERE data.

In the past, ERE has found that this method of estimating tends to underestimate the project investment by omitting interconnecting or common facilities and using generally optimistic estimates. Specific checks using unit cost data derived from ERE studies for other locations and feedstocks indicate that the Petro-Canada investments are low for the process units checked:

- Gasification +70%
- Oxygen +80%
- Methanol Synthesis +90%
- MTG +20%

7.2.4 Project Investment Checks

Based on costs from other studies for projects with similar facilities at comparably remote locations, the Petro-Canada estimate appears low by a substantial amount. These studies are for plants with higher capacities, using different coal feedstocks and located either in the United States or Australia. Adjusting to the same capacity, location and time frame as the Petro-Canada estimate gives the following relative investments, which include Owner's Costs and Infrastructure.

	Location	Gasification Technology	Investment Relative to Petro-Canada Estimate	
			Constant Coal Feed	Constant Gasoline Production
o Petro-Canada	British Columbia	Koppers-Totzek	Base	Base
o Australian Study	Wandoan	Shell	+60%	+120%
o U.S. Study	Wyoming	Shell	+160%	+180%
o U.S. Study	Wyoming	Lurgi	+100%	+140%
o DOE	Wyoming	Koppers-Totzek	+100%	+110%

The significantly lower level of the Petro-Canada estimate is attributable to understatement of the process unit costs cited earlier and specific omissions and deficiencies discussed in Section 7.2.5.

TABLE 7.1

General Facilities

- Common facilities (i.e., roads, paving, sewers, piperacks, area lighting, etc.)
- Interconnecting process and utility lines
- Process/instrument air supply
- Safety/flare systems
- Tankage other than specified
- Fire protection
- Electrical distribution
- All buildings and fittings
- Parking lots
- Site preparation
- Product loading (i.e., gasoline, LPG and sulphur)
- Rail spur
- Communications (within the complex)
- Fencing and landscaping

7.2.5 Estimate Omissions/Deficiencies

Engineering management services for the overall project do not appear to be included. The turn-key costs obtained by Petro-Canada include the engineering and construction management only for that particular unit. The planning, basic design, project management and start up services normally provided by an integrated ERE/Owner Team for the overall project would typically add about 10% to the capital investment for a project of this size.

The estimate provision for offsite facilities is too low. The estimate includes specific sums for steam generation, cooling towers, water treatment and storage. Petro-Canada also includes an allowance at 10% of the total installed cost to cover General Facilities (see Table 7.1). The total amount included for Common Facilities, Offsites and Utilities is 24% of the onsites investment. The four studies mentioned earlier include 34% to 85% of onsites for these facilities.

Execution of this project will require additional Owner's Costs including pre-start up personnel, personnel related costs, project planning/administration, headquarters charges, taxes, permits, environmental studies/EIS, community contributions, expensed construction, consultants, site investigations, etc. Historically, these costs represent 10 to 15% of capital investment.

Based on similar studies for projects in the United States and Australia, a considerable investment (10 to 20% of project) may be required for infrastructure for this project. The review identified several potential areas not covered in the estimate, e.g.,

- permanent housing
- roads and bridges
- water supply
- solid waste disposal

TABLE 7.2

Proration Slopes

ERE recommends the following proration slopes to bring the Petro-Canada units down to the required capacity:

- Coal Preparation	- 0.7
- Air Separation	- 1.0*
- Gasification	- 1.0* (i.e., divide by 13 and multiply by 7 to include spare)
- Raw Gas Compression	- 1.0*
- CO Shift	- 1.0*
- RECTISOL	- 1.0*
- Sulphur Recovery (Stretford)	- 0.3
- Methanol Synthesis	- 1.0*
- Hydrogen Recovery (PSA)	- 0.8 (per Union Carbide advice)
- Methanol to Gasoline (MTG)	- 0.8 (per Mobil advice)
- Refrigeration	- 0.5
- Steam Generation	- 0.6
- Cooling Towers	- 0.9
- Water Treatment	- 0.4
- Tankage	- 0.9

* Multi-train units divisible by 2

The contingency allowance is inappropriate. The basis for the estimate is in the screening stage of project development. Normally a 25% to 30% project contingency is included for this type of estimate rather than Petro-Canada's 15%.

7.2.6 Adjustments to Half Size Plant

The Petro-Canada estimate was for a project using 9600 dry tonnes per stream day of coal refuse. To develop an investment cost for a half-size plant, the following approach is used:

- Prorate the Phase IA unit costs down to half capacity using slopes provided in Table 7.2.

- Add overall conditioning in the range of 100 to 150% across the board to all units, including offsites and utilities. This is intended to cover scope completion, overall project services, estimate methodology, location factors and additional project contingency.

- Add Owner's Costs at between 10 to 15% and add Infrastructure at between 10 to 20% of the process plant TEC, subject to an assessment of the specific requirements.

- Test the economics for investment sensitivity.

TABLE 7.3
Conversion Complex Estimate Breakdown
M$-1Q83

	Phase IA, TEC(1)	Half-Size Plant, TEC Phase IA(2)	Half-Size Plant, TEC Current(3)
ONSITES			
Coal Preparation	46	29	39
Air Separation	184	92	142
Gasification	277	148	217
Raw Gas Compression	61	31	35
CO Shift	69	35	38
RECTISOL Acid Gas Removal	72	36	40
Stretford Sulphur Plant	10	8	9
Methanol Synthesis	110	55	91
Hydrogen PSA Unit	9	6	6
MTG Plant	210	120	124
Subtotal	1,048	560	740
UTILITIES			
Refrigeration	13	9	
Boiler/Superheater	11	8	
Cooling Water System	14	7	
Water Treatment	86	66	
Subtotal	124	90	150
OFFSITES			
Storage/Blending	23	12	
General Facilities	120	58	
Subtotal	143	70	280
	=====	=====	=====
TOTAL Plant Costs	1,315	720	1,170

Notes: (1) Total plant cost of original Phase IA plant at 9,600 t/sd (m.f. including 15% contingency.

(2) Total plant cost of original Phase IA plant at 4,800 t/sd (m.f including 15% contingency.

(3) Total plant cost of current conversion complex at 4,800 t/sd (m.f including 30% contingency.

7.3 Feedstock Preparation Plant Estimate

The capital cost estimates for the feedstock preparation plant and for the installations upstream of the complex including the dump station, slurrying take-offs from the Elkview Plant, the slurrying and pumping stations, the slurry line and river crossing, etc. are all based on 1984 cost estimate data. This data is generated based on vendor quotes for appropriate equipment and the installation cost for the equipment and related works or based on in-house construction cost information available to Westar Engineering Ltd. that is pro-rated for this application. A fair amount of design and sizing of major constituent elements is undertaken at this phase in order to arrive at reasonably reliable forecast costs. In similar pre-feasibility studies the figures generated are considered to be accurate to $\pm$ 25%. All cost figures developed for this portion of work are specific to site E in Sparwood, are inclusive of all equipment, labour and materials and also include Provincial taxes, Federal taxes, and duty where applicable.

7.4 Conversion Complex Estimate

Based on the criteria set out in Section 7.2.6 the unit costs from the Phase IA report are prorated based on the factors recommended by ERE and shown in Table 7.2. The Phase IA General Facilities estimate is not included in this proration as it is accounted for in the adjustment factor below. Based on ERE's previous estimating experience (section 7.2.4) a factor of +105% is selected to convert the total installed costs of the Phase IA study to total plant costs with a 30% contingency. The revised Conversion Complex estimate breakdown is shown in Table 7.3 and is compared to the Phase IA full size estimate.

TABLE 7.4

Comparison of Cost Estimates @ Half Size

M$ - 1Q83

	Phase IA	Current	Factor
Onsites	487	570	1.17
Offsites/Utilities	138	330	2.39
TOTAL INSTALLED COST (TIC)	625	900	1.44
Contingency	95 (15% TIC)	270 (30% TIC)	2.84
TOTAL PLANT COST (TPC)	720	1,170	1.63
Owner's Costs	44 (7% TPC)	120 (10% TPC)	2.73
TOTAL ERECTED COST	764	1,290	1.69
Infrastructure	-	100 (8.75% TPC)	-
TOTAL	764	1,390	1.82
Working Capital	13 (2% TIC)	8 (EST)	.62
TOTAL PROJECT COST	777	1,400	1.80

TABLE 7.5

Updated Total Project Costs

2Q84 - M$

Onsites	610
Offsites/Utilities	350
TOTAL INSTALLED COST	960
30% Contingency	300
TOTAL PLANT COST (TPC)	1,260
Owner's Costs (10% TPC)	130
TOTAL ERECTED COST	1,390
Infrastructure (8.75% TPC)	110
Working Capital (Estimated)	9
TOTAL PROJECT COSTS	1,510

7.5 Total Project Capital Costs

In addition to the total plant costs of 1,170 M$ (1Q83) estimated in Sections 7.3 and 7.4, there are additional associated capital costs that are referred to as Owner's Costs and Infrastructure Costs.

Owner's costs represent the one-time capitalized expenses associated with the initial startup of the plant. These costs include plant staff buildup, training and associated manpower costs, operating costs associated with "running-in" the equipment, initial catalyst and chemical charges, etc. These costs are estimated at 10% of the total plant costs.

Infrastructure costs typically represent the "off plot" costs associated with the project. These costs include supply of remote services such as water, natural gas and power supply, etc. These costs are estimated at 8.75% of the total plant costs versus traditional levels of about 10% to 20% due to the more developed site and proximity of utilities supply sources.

As shown in Table 7.4 these assumptions result in a total project cost of 1,400 M$ (1Q83) which is about 1.8 times the equivalent Phase IA report cost at an equivalent size. Table 7.5 shows the total project cost updated to 2Q84 dollars.

TABLE 8.1

Summary of Annual Operating Costs

- BASIS: Millions of Canadian Dollars (1984)
  330 days of operation

- FIXED COSTS

Wage Employees	15.0
Admin./Technical	3.1
Supervision	4.3
Insurance/Materials/Taxes	40.2
Prep. Plant Maintenance	1.0
TOTAL FIXED COSTS	63.6

- VARIABLE COSTS

Chemicals/Catalysts	5.8
Coal Feedstock	10.3
Waste Disposal	1.3
Electricity	25.5
TOTAL VARIABLE COSTS	42.9
TOTAL ANNUAL OPERATING COSTS	106.5

8.0 OPERATING COST ESTIMATES

8.1 General

The basis for operating cost estimates is described in this section. Where appropriate, costs are developed by halving the Phase IA cost estimates, e.g., coal feed, power, catalyst/chemicals. A summary of operating cost estimates is presented on Table 8.1.

8.2 Utilities

Power demand is one-half of the Phase IA estimate of net imported power requirement. This is further adjusted to reflect a more accurate estimate of power demand for the refuse preparation plant. The net power requirement for the conversion complex and preparation plant using Koppers-Totzek technology is 123 MW.

Power cost is a function of demand charge and energy charge. The energy charge is 2.16 ¢/KWH and when combined with a demand charge based on 100,000 KVA peak consumption, the overall power cost is 2.64 ¢/KWH (1984).

8.3 Labour Requirements

The total labour force is comprised of process, mechanical, technical and administrative staff. Each of these functional areas feature a mix of wage and salaried employees. The process work force is developed from industry guidelines for the individual process components of the complex. These are summarized on Table 8.2. The total process work force is estimated assuming five workers to cover each shift position and allowing roughly one salaried person for each four wage earners.

The other work areas are estimated as follows:

- mechanical - 0.5 workers per $M of investment (excluding owner's cost, infrastructure and contingency)

TABLE 8.2

Estimate of Process Work Force (Wage Earners)

Process Component	Shift Positions	
	Operator	Controller
Coal gasifier	4	1
CO shift	2	
Acid gas removal	1	1
Sulphur recovery	1	
Methanol synthesis	1	
PSA hydrogen recovery	1	1
Mobil MTG	3	
Gasoline hydrotreater	2	1
Air separation	4	1
Product blending/storage	1	
Water treatment	1	1
Utilities	4	
	25	6

TABLE 8.3

Summary of Total Labour Requirement

	Wage	Salary
Process	155	38
Mechanical	179	29
Technical	3	31
Administrative	10	43
Prep. Plant/Waste Handling	27	5
TOTAL	374	146

- o technical - 0.1 workers per $M of investment
- o administrative - 11% of total staffing requirement

The work force for the refuse preparation plant and for waste handling are estimated separately.

The total labour requirement is 520 workers (see Table 8.3).

For estimating labour costs it is assumed that wage earners are paid $40K/yr. (1984), administrative and technical staff are paid $42K/yr. (1984) and supervisors earn $60K/yr. (1984). These figures include benefits.

8.4 Feedstock

The coal refuse from the Elkview plant is generally available in sufficient quantity to provide the tonnes required daily for the coal liquefaction complex. This material is priced at $6/tonne (1984). When refuse is not available in sufficient quantity from the Elkview plant, material from the waste stockpile is delivered to the liquefaction complex. Roughly 30% of the material from the waste stockpile will be oversized and returned to the stockpile. Therefore, roughly 1.6 tonnes of waste must be handled to make one tonne of feed available to the liquefaction plant when drawing feed from the waste stockpile. The additional handling cost will add $1.6/tonne to the feedstock cost during this mode of operation. This is estimated to be necessary for one month per year during Elkview plant shutdown periods.

8.5 Catalysts and Chemicals

Annual charges for catalysts and chemicals are estimated to be one-half of the cost used in the Phase IA study after allowing for escalation to 1984 $. The resulting annual cost of $5.8 M (1984) does not include the initial charge of catalyst required for start of operations. Initial inventory of chemicals and catalyst charges are included in the owner's cost allowance.

8.6 Insurance, Materials and Taxes

A total annual charge equal to 3.3% of investment (excluding infrastructure and owner's cost allowance) is estimated to cover the cost of insurance, materials and taxes. Maintenance materials for the preparation plant are estimated separately to cost $1.0 M/year.

8.7 Waste Disposal

Volumes of solid and liquid waste are estimated by halving Phase IA amounts. Disposal costs are $1.00/tonne (1984) for solid waste and $0.40/tonne for aqueous waste.

9.0 MARKETABILITY ASSESSMENT

9.1 General

Esso Resources assessed the marketability of the MTG gasoline and associated LPG by-products from the conversion complex. Esso Petroleum Canada (EPC) provided the 1990's value of the MTG gasoline, while Esso Resources Gas Department provided the 1990's value of the LPG by-products.

9.2 MTG Gasoline

From an octane standpoint, the MTG gasoline is below the likely gasoline pool average expected in the 1990's with the advent of low to no lead gasoline requirements. This naphtha stream thus has no octane blending advantages at 88 octane (R+M/2). In addition, limited experience with the use of synthetic materials to date has indicated quality problems not experienced with natural crude based streams. There is consequently a need to carefully examine the blending characteristics of this naphtha.

In view of the excess capacity for naphtha present in industry, this naphtha stream should be viewed as an alternative refinery feedstock. As such, it should be valued at 105% of D_2S_2 crude or 25 ¢/litre (39.90 $/B) 1984 $ at the refinery gate.

Because of the low volume of naphtha (1.3 km^3/sd) the only viable transportation mode is via tank car to Edmonton. Based on 1984 rates the transportation cost from Sparwood, B.C. to Edmonton, Alta. is estimated to be 2.9 ¢/l (4.60 $/B) 1984$. Thus, the netback of MTG gasoline at the conversion complex plant gate is 22.2 ¢/litre (1984).

9.3 LPG and Other By-Products

Esso Resources' Gas Department reports that the 99 m^3/sd of propane and 182 m^3/sd of butane produced by the project can be moved by tank car to northwest U.S. markets. More specifically, the propane is shipped to southern Montana and the butane to northwest Washington.

The propane product requires some content changes to meet U.S. propane specifications of 2 LV% maximum ethane, 5 LV% maximum propylene and 95 LV% minimum propane. With a 0.5 ¢/l penalty, propane at 90-95 LV% purity is exportable to the U.S.

The butane quality is good with the high butanes and iC_4 content commanding a $10/m^3$ premium over field butane.

The propane and butane netback values are therefore estimated at 126 $/m^3$ and 201 $/m^3$ (1984$), or 52% and 83% of D_2S_2 crude, respectively.

The only other by-product is elemental sulphur which is valued at current market prices of 165 $/t (1984$). It is assumed to be stored in proximity to the plant site in block storage.

TABLE 10.1

ECONOMIC BASIS AND ASSUMPTIONS

$ Basis	Mid-1984 Cdn $
No. of Operating Days	330 Days/Year
Construction Start/Finish	1991/1995
Operating Life	20 Years
Start-up Year	1995
Financing	100% Equity
Federal/Provincial Tax Rate	41%
Exchange Rate	$1 Cdn = $0.78 U.S. in 1985/86
	$1 Cdn = $0.80 U.S. in 1987+
Capital Cost Allowance	Three Year (25/50/25) Write-off
Investment Tax Credit	7%
Inflation Rates	See Section 10.2.1
Contingency	30% of TIC
Working Capital	$8.7 Million
Prepaid Royalties and Start-up Costs	Included in Owner's Costs
Manpower Charges (including benefits)	$40 K/Yr. Wage Employees
	$42 K/Yr. Admin./Tech. Employees
	$60 K/Yr. Supervisors

10.0 ECONOMIC ANALYSIS

10.1 General

This section summarizes the economic analysis of the indirect liquefaction study. The analysis includes:

- o economic basis and assumptions
- o base case economics
- o sensitivity analysis

The objective is to establish the general project economics as well as determine which factors are most important economically. Several contingencies and sensitivities are examined and an optimum case is developed.

10.2 Economic Basis and Assumptions

The basis and assumptions used in developing the project economics are summarized in Table 10.1. Further details are described in the following sections.

10.2.1 Project and Economic Parameters

The evaluation is based on the following parameters:

- o production capacity for 330 operating days per year is 1282 m^3/sd unleaded gasoline, 182 m^3/sd butane and 99 m^3/sd propane.

- o construction begins in 1991 with the project coming on-stream in 1995 at 50% of effective capacity, increasing to 87.5% in 1996, and operating at effective capacity thereafter.

- o operating life of the project is 20 years.

Financial and economic parameters include:

- o project financing with 100% equity.
- o a 41% Federal/Provincial tax rate.
- o project sponsors are in currently taxable positions during the years of capital expenditures.
- o a three year Capital Cost Allowance (CCA) and Investment Tax Credit at a 7% rate.
- o the following escalators are used:

Escalator	Average Impact	Applications
Consumer Price Index	5.7%/yr.	coal refuse feed
Non-Residential Construction Index	5.9%/yr.	capital costs
Average Weekly Wage Index	6.5%/yr.	labour costs
General Operating Cost Index	6.2%/yr.	electricity cost catalysts/chemicals waste disposal insurance, materials, taxes

10.2.2 Capital Costs

Estimation of fixed capital costs is described in Section 7.0 and summarized on Table 7.5.

TABLE 10.2

Summary of Base Case Price Projections

	Saudi Light F.O.B. Persian Gulf $U.S./m^3	Alberta Field Price $CDN/m^3	Plant Netbacks Cents/Litre		
			Gasoline	Propane	Butane
1984	182	239	22.2	12.0	20.1
1985	182	239	22.0	12.5	20.4
1990	234	314	28.8	15.3	23.7
1995	344	459	42.4	26.2	36.7
2000	506	675	62.7	40.7	53.5
2005	744	991	92.6	59.8	78.6
2010	1093	1457	136.8	87.8	115.6
2015	1605	2140	201.9	129.0	169.8

Canadian $ @ $0.78 U.S. in 1985 increasing to $0.80 U.S. by 1987.

The construction period is estimated at five years with the following spending profile:

Year 1	8%
Year 2	22%
Year 3	45%
Year 4	19%
Year 5	6% (start-up year)

Working capital requirements are estimated as follows:

	$ Million (1984)
Receivables, 45 days sales	11.6
Payables, 30 days production @ cost	(6.7)
Inventories, 15 days production @ cost	3.4
Coal feed, 15 days @ cost	0.4
Catalyst/Chemicals	Incl. in fixed capital
Spare parts	Incl. in fixed capital
TOTAL WORKING CAPITAL	$8.7 Million

10.2.3 Selling Prices

Selling prices for gasoline, propane, butane and sulphur are estimated for 1984 based on the marketability assessment as reported in Section 9.0. These are then escalated for the 1984-2015 period following the trend of crude price escalation set by the Canadian Producers Association forecast of December 27, 1984. This calls for declining real crude prices in 1985 and 1986, no real growth in 1987 and 1988 followed by 2% per year real growth in prices after 1988.

A summary of the price projections for crude oil, unleaded gasoline and propane/butane are shown in Table 10.2.

FIGURE 10.1

OPERATING COST COMPONENTS

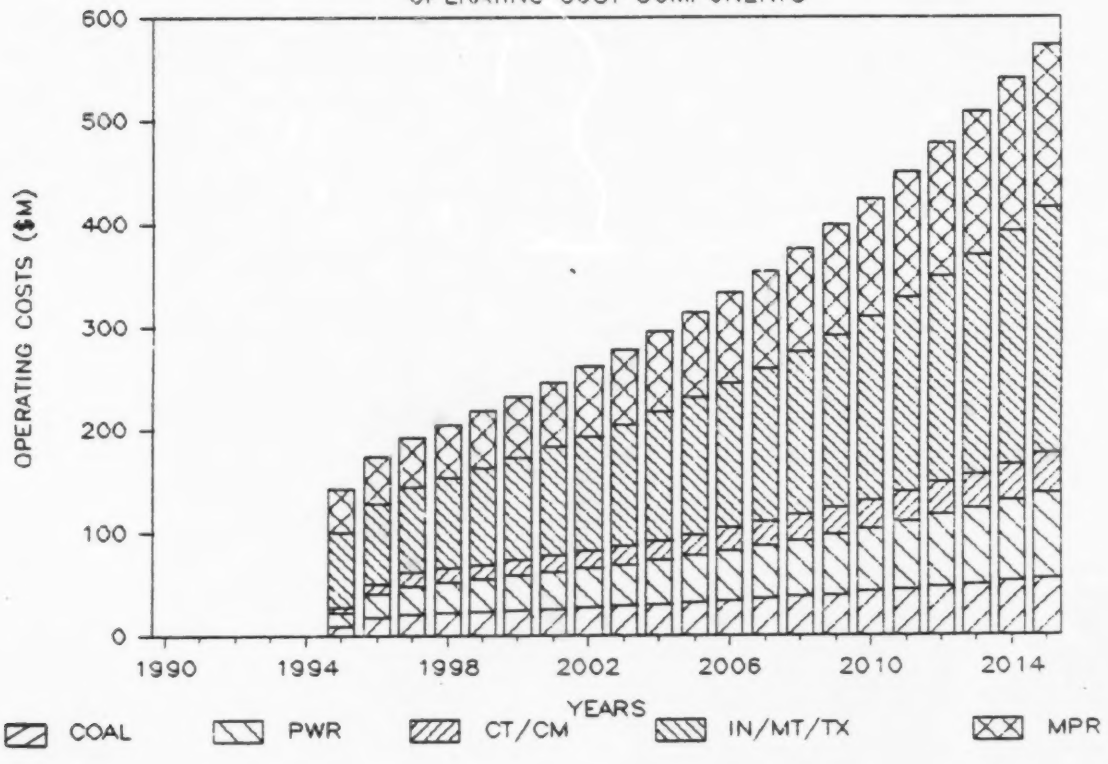
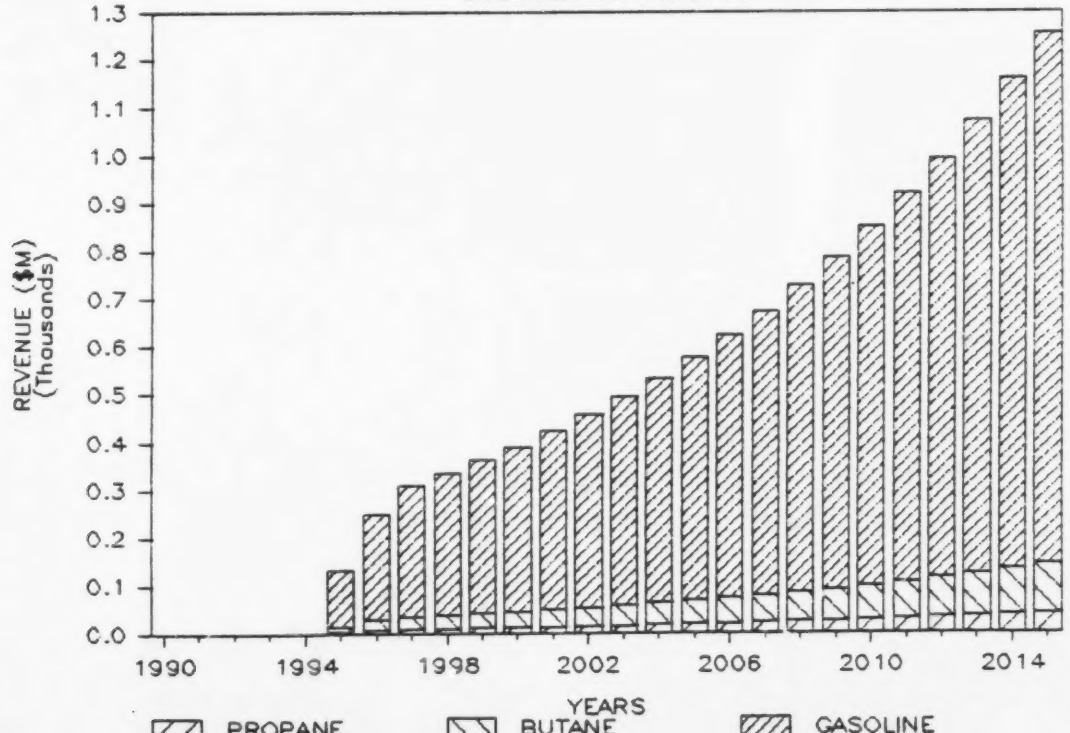


FIGURE 10.2

SOURCES OF REVENUE



10.2.4 Manufacturing Costs

The basis for manufacturing costs is given in Section 8.0 and summarized on Table 8.1. The factors used to escalate the manufacturing cost components throughout the project life are described in Section 10.2.1. Figure 10.1 shows the relative magnitude of operating cost components.

Process royalties are capitalized and included in the capital cost estimate for the facility. Start-up expenses are included in the owner's cost allowance.

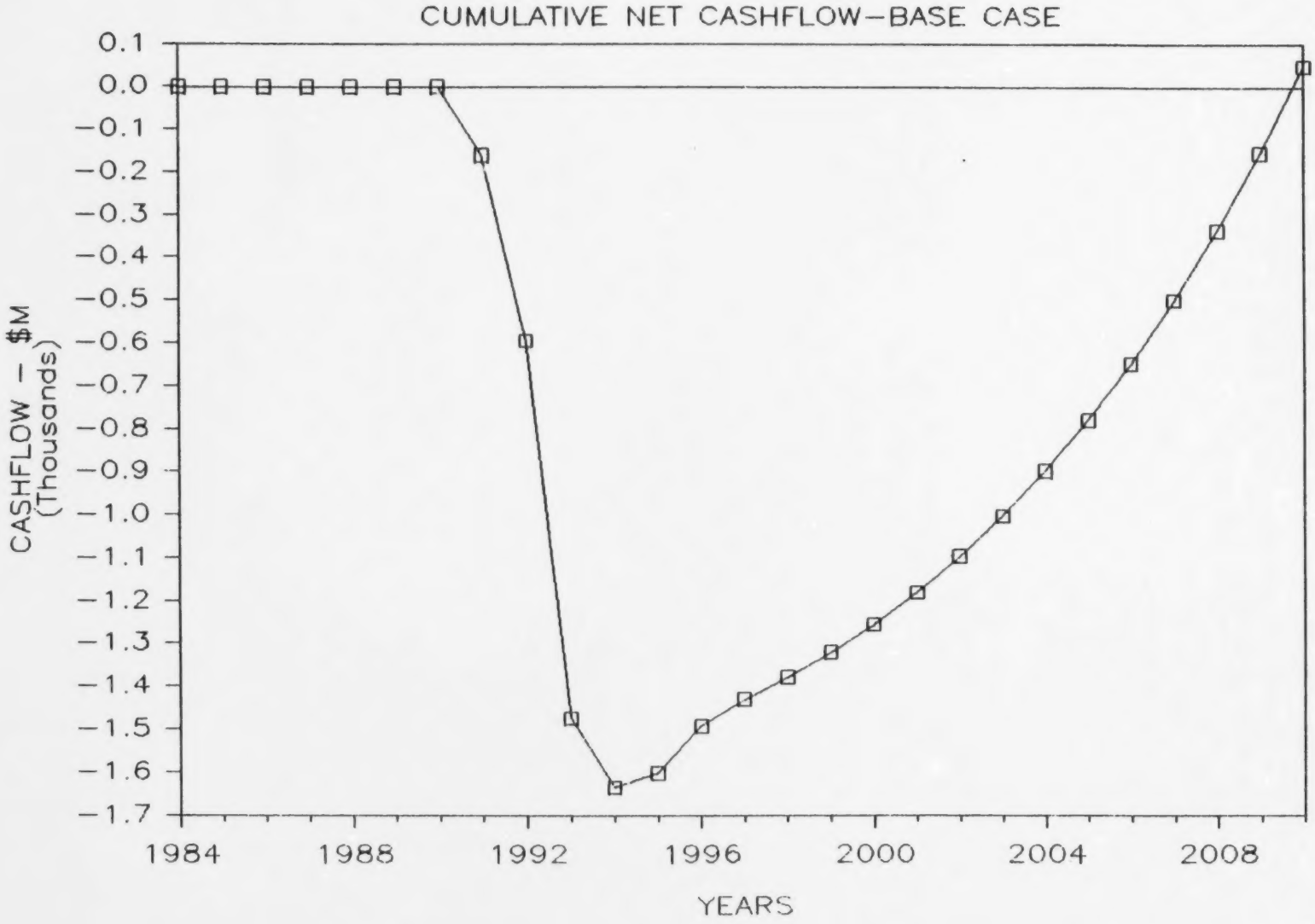
10.2.5 Government Support

Synfuels projects in Canada and the U.S. typically receive financial support to promote economic viability and protect against the highly volatile and unpredictable energy pricing/policy environment. These arrangements are not intended to guarantee a project return and therefore "cost-plus" type deals are generally not acceptable. Loan or price guarantees and tax relief or adjustment are the preferred options.

In this study two types of government support are introduced. Some tax relief is assumed for the base case and price guarantees are assumed in the sensitivity analysis.

As the principal product is gasoline, and most project revenue comes from the sale of gasoline (see Figure 10.2), it is possible that money from the Federal and B.C. Provincial Government's gasoline taxes could be applied to the project. This is assumed to occur for the life of the project. Federal sales tax is collected in an amount equal to 10% of the manufacturer's selling price (wholesale). B.C. provincial road tax is collected in an amount equal to 20% of the retail selling price. Therefore, a total supplement equal to 30% of the project gasoline netback is added to project

FIGURE 10.3


CUMULATIVE NET CASHFLOW—BASE CASE
CASHFLOW - $M
(Thousands)
0.1
0.0
−0.1
−0.2
−0.3
−0.4
−0.5
−0.6
−0.7
−0.8
−0.9
−1.0
−1.1
−1.2
−1.3
−1.4
−1.5
−1.6
−1.7
1984
1988
1992
1996
2000
2004
2008
YEARS

revenues. In other words gasoline netbacks are increased by 30%. The additional revenue is subject to taxation in the context of normal project tax calculations. This tax relief is included in the base case economics.

10.3 Base Case Analysis

The economics calculated on the base case parameters noted in Section 10.2 are:

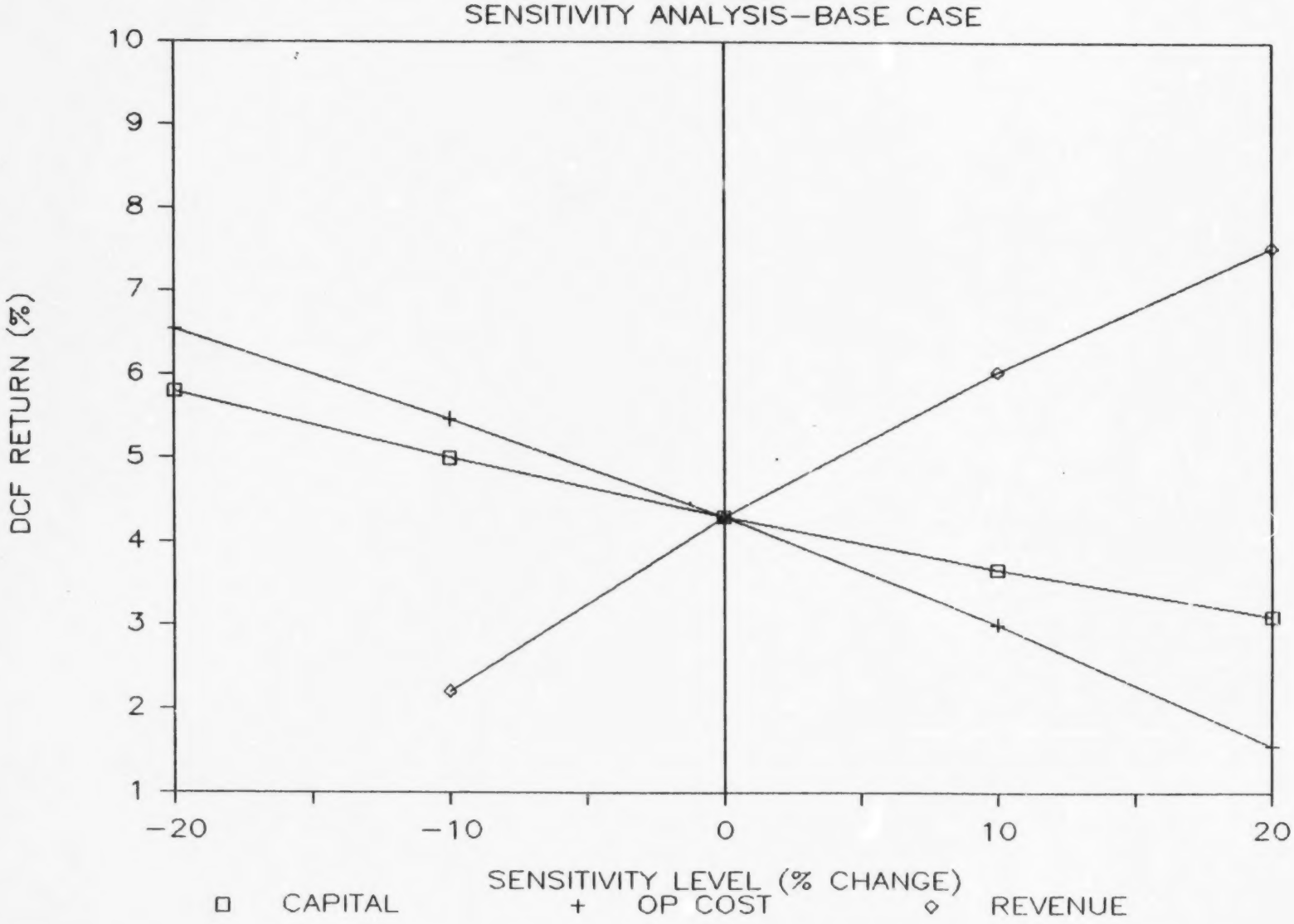
DCF RATE-OF-RETURN	4.3%
PAYBACK PERIOD (EXCLUDING CONSTRUCTION PERIOD)	16 years

At 4.3% the base case DCF return is not good enough for the project to be seriously considered. In fact, the inflation adjusted or "real" rate of return would be zero for this project. With cost of capital funds in the 8-10% range no prudent investor would look to this project for economic gain.

The cumulative net cash flow profile shown in Figure 10.3 indicates a maximum cash impairment of $1.6 billion at the end of the construction period and a project payback of 21 years from start of construction. This is an extremely long period of time even by comparison with other capital intensive synfuels projects. The risk exposure over such a long period is another major stumbling block for the project.

The three principal components of operating cost are insurance/materials/ taxes at 38% of the total, followed by power charges at 24% and labour at 21%. Total variable costs are roughly 40% of the cash manufacturing costs. The high burden of fixed costs is to be expected with the relatively small production scale of this facility. Coal feed comprises less than 10% of total costs.

FIGURE 10.4


SENSITIVITY ANALYSIS—BASE CASE
DCF RETURN (%)
10
9
8
7
6
5
4
3
2
1
−20
−10
0
10
20
SENSITIVITY LEVEL (% CHANGE)
CAPITAL
OP COST
REVENUE

10.4 Sensitivity Analysis

The following sensitivities are examined:

- general sensitivity to revenue, capital cost and operating cost changes
- a Texaco gasifier case which features lower capital cost and lower operating cost
- an optimum case which incorporates the Texaco gasifier and government price guarantees to allow a 15% DCF return.

The basic sensitivities of revenue, capital and operating cost are illustrated in Figure 10.4. The project is noticeably less sensitive to changes in capital cost than operating cost or revenue. Even a 50% rise in capital cost only drops the project return by 2.5 percentage points.

Change In	DCF (Base Case = 4.3%)	
	+20%	-20%
Revenue	7.5	No Return
Capital Cost	3.1	5.8
Operating Cost	1.6	6.5

Unfortunately even large favourable changes to revenue or operating cost do not raise the return significantly above 10% DCF. On the upside, the project return is most responsive to increased revenue. This is useful when considering price guarantees.

The following adjustments are made to the base case economics to reflect the Texaco technology:

- capital costs lower by $111 million through simpler feed preparation plant and no raw gas compressor.

- operating costs lower by $16 million per year due to lower power consumption and less preparation plant maintenance.

FIGURE 10.5

CUMULATIVE NET CASHFLOW—TEXACO CASE

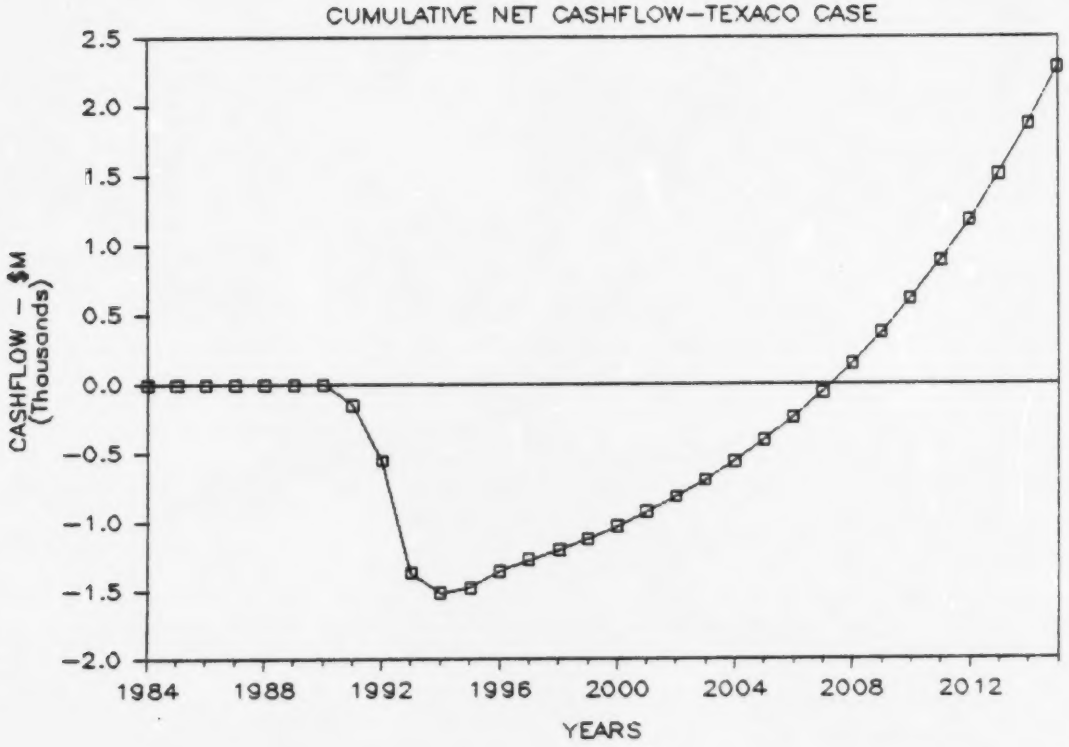
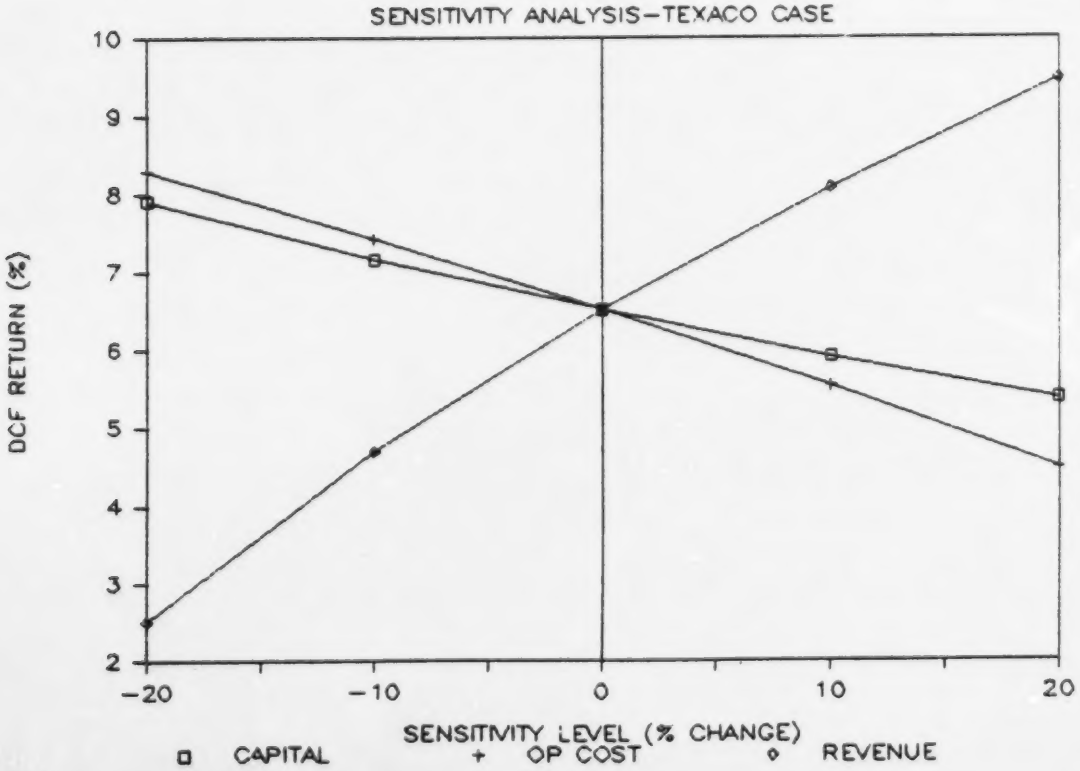


FIGURE 10.6

SENSITIVITY ANALYSIS—TEXACO CASE



With these changes the DCF rate-of-return improves somewhat to 6.5% with a payback period of 14 years from start-up and maximum cash impairment of $1.5 billion at the end of construction. While this return is marginally better than the base case 4.3% return it is still too low for economic viability under normal circumstances. The cumulative net cash flow profile and general sensitivity analysis results for the Texaco case are presented in Figure 10.5 and Figure 10.6.

Finally, an optimum case is developed which incorporates a price guarantee. This guaranteed price is set well above the market prices projected for the base case economics. The price guarantee needed to achieve 15% rate of return for the Texaco case is a gasoline price of 45 ¢/litre F.O.B. Edmonton refinery in 1984. This price escalates in step with the crude price forecast. The difference between the guaranteed price and the forecast gasoline price in any given year becomes a source of additional revenue to the project. As such it enters into the tax calculation as additional profit and taxes are paid accordingly. The cumulative value of this subsidy is $2.35 billion (1984$). Roughly speaking the price guarantee doubles the price received by the project for gasoline sales. The amount of subsidy is also roughly equal to the price of a barrel of crude. Therefore, doubling the crude prices assumed for project economics would accomplish the same thing.

10.5 Alternative Products

The following cases are examined to determine if the project economics can be improved by selling product produced at intermediate stages.

- production and sale of chemical synthesis gas (CSG)
- production and sale of Methanol

Based on ERE guidelines the following economic adjustments are made for each case relative to the base case making gasoline.

TABLE 10.3

SUMMARY OF METHANOL AND NATURAL GAS
PRICE PROJECTIONS

	NATURAL GAS ALBERTA PLANT GATE $/GJOULE	METHANOL F.O.B. EDMONTON ¢/LITRE
1984	2.55	14.7
1985	2.55	15.0
1990	4.05	20.9
1995	6.90	31.0
2000	10.20	42.8
2005	14.95	59.7
2010	21.95	83.6
2015	32.25	118.1

	Methanol	CSG
Product Yield (kjoule basis)	+6%	+22%
Investment (TIC)	-12%	-26%
Power Cost	-15%	-32%
Labour Cost	-10%	-19%
Catalyst/Chemicals Cost	-16%	-38%

To estimate revenue for these cases, the CSG is assumed to sell at joule-parity with export natural gas (using the CPA natural gas price forecast) and methanol is assumed to sell for 14.7 cents per litre F.O.B. Edmonton, escalating with natural gas prices and CPI. The methanol price forecast is provided by Celanese Canada Inc. These price projections are summarized on Table 10.3.

The economics calculated for these cases are compared with base case returns as shown below.

	DCF RATE-OF-RETURN
Base Case	4.3%
Texaco Case	6.5%
CSG Case	0.2%
Methanol Case	5.7%

The CSG and Methanol cases do not include any government support whatsoever. If government support is added to the extent that it is included in the base case and Texaco cases, then the CSG and Methanol case returns climb to 6.3% and 9.4% respectively. While these returns are at least as good or better than other sensitivity cases a 15% hurdle rate for economic viability is still not achieved. Even with the Methanol case total government subsidy of $1.8 billion (1984) is still required to achieve a 15% DCF rate of return.

10.6 Upgraded Feed

If it is possible to lower the ash content of the coal refuse by introducing a further beneficiation step, then the cost of building the gasifier and air separation sections of the complex can be reduced. ERE guidelines indicate that a 2.5% reduction in total project cost is possible with a 50% reduction in feed ash content. At this level of saving a beneficiation step is not economically attractive.

11.0 CONCLUSIONS

11.1 Gasifier Process Selection

A review and critique of the gasification process evaluation and comparison concludes that there is no advantage for the Koppers-Totzek process over the Texaco process.

Progress made in Texaco process development and commercialization over the past two years has been sufficient to close the gap that previously existed when the two processes were compared in the Phase IA report. The most significant developments are as follows:

- o first commercial Texaco unit started up in 1983 and three commercial scale units are currently operating
- o with either K-T or Texaco gasifiers, high service factor is obtained by providing a spare gasifier
- o technical risks related to slurry pumps, abrasion in reactor burners, and operability of the heat recovery system are currently judged to be no worse than K-T performance concerns

The overall gasifier comparison using process, technical risk and commercial factors shows that no significant difference exists between Texaco and K-T processes for gasification.

Two independent studies show an economic advantage for Texaco in terms of operating cost and investment savings when gasifying bituminous coals. Therefore, with other factors considered equal, the Texaco process looks better on the strength of economic performance expectations.

11.2 Downsizing to Half-Scale

A careful analysis of the Phase IA capital cost estimate shows that it is understated by a factor of two to three. Major reasons for this are:

- o methodology used results in underestimated basic facility cost
- o estimate does not cover the full scope of facilities and services needed
- o project contingency level is too low for the stage of project development
- o owner's costs are incomplete and the infrastructure investment is excluded.

Following a revised estimating procedure a total project cost of $1.5 billion (2Q84) is determined. This is about 1.8 times the equivalent Phase IA report cost at an equivalent size.

11.3 Marketability Assessment

In terms of octane quality the MTG gasoline is below the likely gasoline pool average expected in the 1990's with the introduction of low to no lead gasoline requirements. Recognizing potential quality problems due to the synthetic nature of the material and in view of the excess capacity for gasoline-range naphtha in the industry, the MTG gasoline is valued at 105% of D_2S_2 crude (or $\$251/m^3$) at the refinery gate.

The propane and butane products are saleable in the northwest U.S. markets at netbacks equal to 52% and 83% of D_2S_2 crude respectively.

11.4 Economic Viability

With higher capital costs, lower product netbacks and a half-scale plant, it is not surprising that economic returns are considerably poorer than those reported in the Phase IA study. Analysis of the "Base Case" economics indicates an after tax DCF rate of return of 4.3% with a 16-year payback period (excluding four years of construction). This is not good enough for the project to proceed without some form of government assistance. Changes in major parameters (selling prices/capital costs/operating costs) of ±20% generally leave the economics in the 2 to 8% DCF range. The project return is, however, most sensitive to increased revenue such as might be recovered from government price guarantees.

Price guarantees are applied to a revised case which incorporates Texaco gasification technology. The lower capital and operating costs raise the base case return to 6.5% DCF. Using this case, a price guarantee of $\$453/m^3$ for gasoline F.O.B. Edmonton refinery is needed to achieve 15% DCF rate of return for the project. Given the price forecast assumed for the base case return calculation the government subsidy is $2.35 billion (1984$) over the project life. It is only with this level of support that the project is economically viable.

Alternative project configurations where chemical synthesis gas or methanol are the prime products do not exhibit significantly improved economics. Best returns are achieved with methanol product, but even then government subsidy of $1.8 billion (1984$) is needed over the project life to achieve 15% DCF rate of return.

Westar Mining Ltd. have previously examined the possibility of upgrading the refuse feed by further ash removal using agglomeration, grinding for better coal release, coal/water processing, etc. All of these tests yielded a low recovery at moderate cost. This information combined with ERE analysis of potential cost savings for better feed quality lead to the conclusion that upgraded feed will not improve the economic viability of the project.

12.0 RECOMMENDATIONS

Based on the unattractive base case economic results it is recommended that no further work be performed on this project. This recommendation is further confirmed by the poor economics that result when alternative process configurations are considered as upside sensitivities.

The project should be re-initiated if dramatic increases in crude price are predicted or if prospects are good for a large measure of government subsidy.

13.0 ATTACHMENTS

Attachment I

TENNESSEE VALLEY AUTHORITY
CHATTANOOGA TENNESSEE

CONCEPTUAL DESIGNS & ASSESSMENTS OF A COAL GASIFICATION DEMONSTRATION PLANT

VOLUME I - SUMMARY

PROJECT 5672-N

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C F BRAUN & CO

Alhambra, California

October, 1980

EXECUTIVE SUMMARY

INTRODUCTION

This is C F Braun & Co's final report concerning conceptual designs and assessments in support of Phase I of TVA's Coal Gasification commercial demonstration plant. The report is organized into five volumes. Volume I is a nonproprietary document and provides a summary of the major technical and financial aspects of all three of the gasification processes assigned to us on this project. It is intended to provide an overview of all of the work accomplished without reference to the specific gasifier reports.

Volumes II, III and IV provide detailed data on the Koppers-Totzek, Texaco and Babcock & Wilcox gasification processes respectively. Each volume contains nonproprietary information pertaining to the specific process documented.

Volume V contains the task reports prepared for this project concerning process selection studies and plant studies along with TVA's Design Criteria document that provided the basis for our work. Also included in this volume is the Gas Cost Guidelines writing and the Building Sketches, that are common to all three processes.

SCOPE

The scope of our work presented in this report consisted of the preparation and evaluation of the conceptual designs of commercial demonstration plants for three coal gasification processes. Each process produces medium Btu gas from a feed of approximately 20,000 tons per day of coal. The processes assigned to Braun were Koppers-Totzek, Texaco and Babcock & Wilcox. The conceptual design for each of the gasification processes consisted of the preparation of the following information and data.

- Process selection studies for the various process subsystems associated with the gasification unit.
- Plant studies concerning the overall facility.
- Engineering designs for the three gasification processes along with their selected process subsystems and associated facilities, complete with material and energy balances.

EXECUTIVE SUMMARY

SCOPE Continued -

- o Environmental assessments
- o Site plans and equipment arrangement drawings.
- o Capital cost estimates, gas costs and sensitivity analyses.
- o Project schedule for the design and construction of the commercial demonstration plant.

FINANCIAL

Conceptual commercial process designs were prepared for the Koppers-Totzek, Texaco and Babcock & Wilcox processes in support of Phase I of the TVA's medium-Btu coal gasification demonstration project. Capital and operating costs were estimated based on a Kentucky #9 coal feedstock. All three designs were based on a consistent set of design criteria, economic ground rules, and estimating procedures.

The total plant investment and the annual operating cost in 1980 dollars are shown below. The accuracy of the estimate is +30 to -15 percent for each process on an absolute basis. When comparing processes, the relative cost differences are more accurate. It must be emphasized that these results are highly dependent on the assumptions and data base used in the preparation of the conceptual designs.

ASSUMPTIONS The assumptions are presented in the Appendix, Volume V, in the section that presents the TVA Design Criteria.

DATA BASE The data used to generate the process designs varied among the processes. The Koppers-Totzek process design was based substantially on commercial experience. The Texaco process design was based on pilot plant level work. The Babcock & Wilcox process design was based on computerized projections and extrapolations of pilot plant work done twenty years ago and B&W's extensive experience with coal-fired boilers.

EXECUTIVE SUMMARY

FINANCIAL Continued

PROCESS	TOTAL PLANT INVESTMENT MILLIONS OF DOLLARS	ANNUAL OPERATING COST MILLIONS OF DOLLARS
Koppers-Totzek	2,000	430
Texaco	1,900	340
Babcock & Wilcox	2,150	385

DESIGN RELIABILITY

The reliability of the estimate is affected by the reliability of the design. For an unproven design there is always the possibility that additional equipment may be needed, a bottleneck may develop, or performance may not be as efficient as envisioned in a conceptual design. In this respect, the Koppers-Totzek process may be at a disadvantage in the economic comparisons. Krupp-Koppers is already well along the learning curve and has provided equipment to solve previously unanticipated technical problems. An example of this is the NO_x converter. Texaco and Babcock & Wilcox have yet to learn the problems encountered in a full scale plant. It is possible that a commercial demonstration plant based on either of these two processes may require additional capital expense.

EXECUTIVE SUMMARY

DESIGN CRITIQUE

The three designs are similar except for the gasifiers. Therefore, an evaluation of the advantages and disadvantages of each design reduces to an evaluation of the three gasifier subsystems.

KOPPERS-TOTZEK The major advantage of the K-T process is the fact that it is a commercially demonstrated process. The major disadvantage is its lower thermal efficiency relative to second generation processes. This lower efficiency results from several factors. Atmospheric pressure gasification requires a large energy input into the compression step. And the K-T process does not recycle the unreacted carbon. This represents about a five percent loss of available heating value contained in the coal feed.

An additional disadvantage of the K-T process is that its environmental impact may be greater, since it produces a larger volume of ash. The total ash that must be disposed of as fly ash, amounts to about seventy-five weight percent. This is much higher than for the other two processes. Fly ash is a more undesirable material than slag because of it's leachability. Slag is also a more saleable byproduct.

TEXACO The primary advantages of the Texaco process are (1) Operation at the desired product pressure, (2) Complete recycle of unreacted carbon, and (3) Production of all ash in the slag form.

While operation at pressure is a theoretical advantage in a conceptual design, it could prove to be an operational disadvantage in practice. Visual inspection of the gasifier interior is no longer possible. Also bringing the coal to reactor pressure and letting down the ash are more difficult operations than compressing the product gas.

One of the disadvantages of the Texaco process is the lack of commercial operation. However, the reactor was successfully scaled up by a factor of ten when the pilot plant in Oberhausen, West Germany was designed. This lends encouragement towards the success of another factor of ten scale-up to achieve the capacity of the unit designed for TVA.

EXECUTIVE SUMMARY

DESIGN CRITIQUE Continued

BABCOCK & WILCOX The primary advantage of the B&W design is its high thermal efficiency resulting from (1) High carbon conversion achieved by fly ash, and (2) High pressure steam superheat by indirect exchange with the product raw gas stream.

The high pressure steam superheat is unique to the B&W design. It obviates the need for a fired superheater using product gas for fuel, as provided in the K-T and Texaco designs. The B&W steam superheater saves about 3.3 percent of the product gas heating value.

The main disadvantage of the B&W process is the lack of integrated pilot plant operation. Hence, the data base available for scale up to commercial size is not as large as for the Texaco process. This subject is discussed more fully in the next section under Technical Evaluation of Risk.

SUMMARY

We believe that the three conceptual designs developed for this project and detailed in accompanying volumes have several unique features. Examples are the zero liquid effluent and the fact that a coal fired boiler is not required.

Because of the configuration of the site and its limited access, special designs will be required to impound the ash and slag, and special procedures will be needed to circumvent the problems imposed by the site during the construction phase. Such items are addressed in this report.

From the Phase [illegible] design procurement and construction schedule standpoint, on a fast track basis, it appears possible that all three processes could come close to meeting TVA's target completion date.

ECONOMIC SUMMARY

ESTIMATE DESCRIPTION

The estimates for the three processes are factored estimates. Working from equipment takeoffs, the engineered equipment costs are obtained from supplier rough price quotations, by adjusting prices from quotations of similar equipment, and in-house estimates. Then installation factors for each type of equipment are applied. The installation factors include the cost of all bulk materials - instruments, foundations, structural steel, piping, electrical, insulation, and painting. They also include all field labor, other proratable labor and construction costs. The installation factors reflect current labor rates and conditions in the Northern Alabama area. A percentage allowance for home office costs was added to all the above.

The home office cost covers project management, engineering, purchasing, estimating, cost control, scheduling, accounting, and other home office and consulting costs.

A contingency allowance is included to cover costs of unforseen and unknown items required in the scope of work. In addition, the contingency is intended to cover possible shortcomings in the methods used to prepare the estimates.

The contingency does not include any money for changes in the scope of work.

Braun was unable to estimate the engineered equipment costs for the oxygen plants, the gasification units, the sulfur removal units or the sulfur prilling units. Vendor quotes were used for these units. Adjustments were made for in-house alterations and to compensate for any exceptions taken by the vendors. Home office costs and a contingency were added.

Braun would expect the overall reliability of the estimates to be plus 30 percent to minus 15 percent.

Table 1 summarizes the installed capital cost by subsystem for the three processes. The estimate is in January 1980 dollars. Also included are the disbursement tables for the three processes. They indicate the actual cash disbursement for each cost center. Each cost center is broken down into capital cost, O&M and working capital by year, along with the yearly totals.

ECONOMIC SUMMARY

ESTIMATE DESCRIPTION Continued

Two types of disbursement tables are included for each process. Tables 3 through 5 give the cash disbursements in escalated dollars. Tables 6 through 8 give the cash disbursements with all values expressed as January, 1980 dollars.

Tennessee Valley Authority, Project 5672-N, October, 1980

ECONOMIC SUMMARY

TABLE 1
CAPITAL COST SUMMARY
JANUARY, 1980 DOLLARS

	KOPPERS-TOTZEK	TEXACO	BABCOCK & WILCOX
Civil Work	$ 247,800,000	$ 225,400,000	$ 210,900,000
Coal Receiving and Storage	38,400,000	38,400,000	38,400,000
Coal Preparation	130,900,000	48,100,000	49,000,000
Oxygen Plant	334,000,000	392,000,000	361,600,000
Gasification	560,000,000	600,000,000	880,000,000
Particulate Removal and Compression	170,800,000	-0-	80,000,000
Sulfur Removal	129,600,000	160,000,000	132,000,000
Claus Sulfur Recovery	32,800,000	58,400,000	32,800,000
Scot Tail Gas Treating	20,000,000	25,600,000	20,000,000
Slag and Ash Disposal	6,300,000	6,300,000	11,000,000
Sulfur Prilling and Barge Loading	11,800,000	11,800,000	11,800,000
Steam System	55,200,000	89,600,000	36,000,000
Plant Water	67,000,000	63,500,000	61,000,000
Electrical Equipment & Distribution	100,700,000	90,500,000	98,000,000
Interconnecting Piping	96,600,000	99,500,000	113,500,000
Total	$2,001,900,000	$1,909,100,000	$2,136,000,000
Use	*$2,000,000,000	*$1,900,000,000	*$2,150,000,000
*Includes a percentage allowance			
For Home Office Cost and Contractor's Fee	158,000,000	150,000,000	170,000,000
Contingency	262,000,000	250,000,000	280,000,000

Tennessee Valley Authority, Project 5672-N, October, 1980

ECONOMIC SUMMARY

TABLE 2
ANNUAL O&M SUMMARY

MILLIONS OF JANUARY 1980 DOLLARS

	PROCESS		
	K-T	TEXACO	B&W
RAW MATERIALS			
Coal @ $27.45/ton	206.2	191.4	195.4
PURCHASED UTILITIES			
Power @ $24.05/MWH	89.0	21.7	38.0
CATALYST AND CHEMICALS			
Catalyst	0.4	0.4	0.4
Chemicals	1.8	2.2	1.8
LABOR			
Process Operating Labor	5.0	4.2	3.7
Maintenance Labor	42.8	40.3	50.2
Supervision	13.5	12.7	15.3
Benefits	20.1	18.7	22.6
PLANT ADMINISTRATION & GENERAL OVERHEAD	19.2	17.8	21.5
SUPPLIES			
Operating	1.5	1.3	1.1
Maintenance	28.5	26.9	33.5
TVA CORPORATE OVERHEAD	2.2	2.0	1.9
TOTAL ANNUAL OPERATING COST	430.2	339.6	385.4

Tennessee Valley Authority, Project 5672-N, October, 1980

Attachment II

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CHEMICAL TECHNOLOGY DIVISION

LIQUEFACTION TECHNOLOGY ASSESSMENT - PHASE II: INDIRECT LIQUEFACTION OF COAL TO GASOLINE USING TEXACO AND KOPPERS-TOTZEK GASIFIERS

A. R. Irvine
R. M. Wham
J. F. Fisher
R. Salmon
W. C. Ulrich

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OAK RIDGE NATIONAL LABORATORY
Oak Ridge, Tennessee 37831
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LIQUEFACTION TECHNOLOGY ASSESSMENT - PHASE II: INDIRECT LIQUEFACTION OF COAL TO GASOLINE USING TEXACO AND KOPPERS-TOTZEK GASIFIERS

A. R. Irvine, R. M. Wham, J. F. Fisher, R. Salmon, and W. C. Ulrich

ABSTRACT

This report provides a comparison of two entrained-flow gasifiers — the Koppers-Totzek and the Texaco — for use in a plant to convert coal to gasoline by an indirect method. The results of the current study are also compared with those from a previous study that employed a Lurgi gasifier for the same purpose. Koppers-Totzek— and Texaco gasifiers—equipped plants were estimated to be economically competitive for the operational parameters used in this study when a high-water-content western coal was used as feed, but the Texaco gasifier—based plant was indicated to be more economical than the Koppers-Totzek—based plant when eastern bituminous coal was used as feed. A Lurgi gasifier— and methane reformer—equipped plant was calculated to be significantly more economical than either of the plants equipped with entrained-flow gasifiers when processing western coal; however, the combination of uncertainties and potential for process improvement associated with these gasifiers makes this conclusion tenuous.

To date, the Texaco gasifier has only been used to gasify coal at the demonstration plant scale, whereas the Koppers-Totzek gasifier is in commercial use.

1. EXECUTIVE SUMMARY

This study was concerned with the conversion of coal to liquid fuel by an indirect process using two different entrained-flow gasifiers and two types of coal. The gasifiers were the commercially available Koppers-Totzek (K-T) and the Texaco (TX), which has not been commercially demonstrated; an eastern bituminous and a subbituminous western coal comprised the two feedstocks. The four cases are designated KTE, KTW, TXE and TXW, respectively. Both the Imperial Chemical Industries (ICI) methanol synthesis process and Mobil's methanol-to-gasoline (MTG)

conversion process were employed. In all cases, the rate of moisture- and ash-free (maf) coal feed to the gasifiers was 14.5 Gg/d [16,000 tons/d (TPSD)]. A similar study, performed earlier, was based on the use of Lurgi Mark IV gasifiers to generate synthesis gas from western coal. Some of the results of that study, reported in ORNL-5664,[1] are included here to illustrate the variations in overall yields and product costs that occur with changes in gasifier characteristics/performance.

Fluor Engineers and Constructors, Houston Division, and the Oak Ridge National Laboratory (ORNL) were jointly responsible for carrying out the study. Fluor prepared the conceptual designs, cost estimates, and assessments of the technology; while ORNL performed the economic and process analyses and interpreted the results.

1.1 COMPARISON OF GASIFIERS

1.1.1 Via Conceptual Plant Designs and Cost Estimates

The cost and efficiency data for Fluor conceptual plant designs and cost estimates are shown in Table 1.1. The information in this table indicates that relatively small differences in plant cost arise from the selection of gasifier type but that substantial differences in plant thermal efficiency are a function of gasifier characteristics. The product price is closely related to plant thermal efficiency, not because of variation in the cost of feed materials but because reduced thermal efficiency results in a reduced yield of saleable products without decreasing capital costs. This comparison indicates that the Lurgi gasifier will yield the lowest-cost product when processing subbituminous coal and the Texaco gasifier will provide the lowest cost when feeding bituminous coal. The Koppers-Totzek gasifier is indicated to be on a par (in terms of cost of product) with the Texaco gasifier for subbituminous coal but somewhat more expensive when using bituminous coals.

The base-case economic assumptions that were used in calculating the gasoline prices in Table 1.1 are specified in that table. The use of other economic assumptions will change the absolute values of the

Table 1.1. Comparison of indirect liquefaction plants as affected by gasifier and feed coal selection

	Koppers-Totzek gasifier		Texaco gasifier		Lurgi gasifier[b]
	Eastern[a]	Western	Eastern	Western	Western
Capital investment[c]					
Total, 10^9	3.1	3.0	2.8	2.7	2.9
$/(daily GJ product)	12,562	13,196	11,183	13,970	9,881
Operating cost[c]					
Annual, 10^6	419	404	361	337	384
$/[GJ product (90% on-stream)]	5.17	5.41	4.39	5.31	3.98
Gasoline production rate, BPSD[d]	36,320	34,854	35,660	27,625	42,580
SNG production rate, FOE BPSD[d]	5,054	4,325	6,512	5,168	3,078
Gasoline price, $/gal[e] ($/L)	2.08 (0.55)	2.15 (0.57)	1.81 (0.48)	2.27 (0.60)	1.65 (0.43)
Plant thermal efficiency,[f] %	35	34	44	39	48
Relative plant efficiency[g]	74	72	93	81	100

[a]Refers to type of coal; each case subsequently abbreviated as follows: Koppers-Totzek, Eastern (KTE); Koppers-Totzek, Western (KTW); Texaco, Eastern (TXE); Texaco, Western (TXW).

[b]Gasoline production maximized. Values in this table are less favorable than those reported in ORNL-5664 (see Sect. 3.4 for explanation).

[c]Mid-1979 dollars.

[d]Fuel oil equivalent barrel (FOEB) per stream day (BPSD), 6.33 GJ (6 x 10^6 Btu)/FOEB.

[e]Base case assumed: 100% equity financing; 12% annual after-tax rate of return on equity; 1990 price scenario specified by DOE; 25-year operating life; 5-year construction period; 48% federal income tax; 3% state income tax; 2.75% local property tax and insurance; mid-1979 dollars; sum-of-years digits depreciation for 16 years; coal cost of $1/1.055 GJ ($10^6$ Btu); no salvage value.

[f]Based on ratio of higher heating value (HHV) of products to HHV of feed coal.

[g]Lurgi gasifier-based plant is used as reference.

calculated selling price but will have no effect on the relative standing of the several gasifier-synthesis plant combinations. This condition results from a relatively constant relationship between operating cost and capital investment. The base-case economic assumptions yield a price that is substantially higher (by a factor of about 2) than would be calculated by employing the 1982 depreciation schedules, by using 75% debt financing, and by assuming an annual inflation rate of 6% or greater for both product prices and operating costs.

1.1.2 Via Gasifier Energy Distribution

The performance characteristics of the three gasifiers were compared, employing an adjusted energy balance and the gasifier conditions used by Fluor. Table 1.2 shows the energy necessary to produce oxygen and the mechanical/electrical energy demands as inputs on the basis of the quantity of coal required to yield that product. Steam is evaluated in terms of the mechanical energy it could produce when expanded to usual condenser conditions. This method of comparing efficiencies has the strength that it provides verifiable equal treatment of all candidates; on the other hand, one might argue that equivalent utilization of waste heat at comparable cost is not always possible.

This method of comparison provides a similar ranking to that obtained from the conceptual design and cost estimate results; however, the difference between gasifiers is somewhat less pronounced. The reduction in difference is attributable, in part, (1) to the assumption of higher efficiencies for compressors and steam turbine drivers (but not as high as current technology can provide) in the energy balance calculations, and (2) to a uniform utilization of "waste" heat from all gasifiers.

More favorable K-T and TX gasifier operating conditions than the ones used by Fluor are probable; hence, the relative efficiency rankings for K-T and especially TX are subject to significant improvement.

1.2 COMPARISON OF RESULTS WITH THOSE FROM OTHER DESIGNS

Koppers Engineering and Construction prepared a preliminary design and cost estimate for a similar coal conversion plant fed with eastern

Table 1.2. Gasifier energy terms[a] and efficiencies

	Koppers-Totzek		Lurgi with CH_4 reformer	Texaco	
	Eastern	Western		Eastern	Western
Inputs					
Coal	1.000	1.000	1.000	1.000	1.000
Coal drying	0.013	0.072	-	-	-
Oxygen purification	0.122	0.134	0.079	0.126	0.160
Steam[b]	0.045	0.038	0.177	-	-
Gas cleanup	0.052	0.062	negligible	0.007	0.007
Gas compression	0.192	0.218	0.089	0.067	0.084
Other	0.001	0.001	-0.038[c]	-	-
Total	1.425	1.525	1.307	1.200	1.251
Outputs					
$CO-H_2$	0.654	0.682	0.776	0.645	0.497
Misc. hydrocarbons	0.002	0.002	0.085	0.015	0.013
Unburned carbon[d]	(0.064)	(0.043)	(0.009)	(0.038)	(0.043)
Recovered heat equivalent[b]	0.181	0.194	0.085	0.192	0.290
Total	0.837	0.878	0.946	0.852	0.800
Efficiency					
Outputs/Inputs	0.587	0.576	0.724	0.710	0.640

[a]All energy terms are expressed as a fraction of the lower heating value (LHV) of the gasifier feed coal.

[b]Energy equivalent is expressed as the energy in coal that would be required to generate the amount of energy contained in, or derivable from, the streams under consideration.

[c]Tar generated that could be used either in the gasifier or a combustor.

[d]No credit was taken for the heat content of unburned carbon. Partial utilization of the heating value of this material is possible.

bituminous coal. This group translated the cost estimate into the same time frame and plant capacity as were used for the conceptual design. Their calculations showed that a 14.5-Gg/d (16,000-TPSD) plant using different atmospheric-pressure entrained-flow gasifiers [Koppers-Babcock-Wilcox (KBW) instead of Koppers-Totzek] would cost $2.4 x 10^9; the projected thermal efficiency of the plant was 42.6%. The results obtained by this group are summarized in Appendix K. (This appendix is included for informational purposes only and does not constitute an ORNL endorsement of the material contained therein.)

A conceptual design prepared for the Electric Power Research Institute (EPRI) based on the use of a Texaco gasifier and an ICI methanol synthesis unit feeding Illinois No. 6 coal has a reported thermal efficiency of 57.8% for producing methanol. This efficiency would be reduced to about 51% if a methanol-to-gasoline conversion step similar to the steps employed in the Phase II designs were added. The principal reasons for the increased efficiency are a more concentrated gasifier feed and a number of cases of more effective use of energy resources.

1.3 POTENTIAL FOR PROCESS IMPROVEMENT

The conceptual designs are not optimized. Assumptions were made with regard to process conditions, equipment style, etc., in the interest of expediency. Table 1.3 provides a description of six changes in equipment or process performance assumptions (i.e., actions) that could increase the plant thermal efficiency and, in most cases, reduce product cost. The largest effect on plant efficiency and product cost would result from substantial increases in the solids content of the coal-water slurry fed to a Texaco gasifier.

Most of the six potential improvements in performance can be categorized either as or near current technology.

1.4 COST EFFECTS OF ENVIRONMENTAL CONTROLS

An ancillary study was performed to determine the cost of indirect coal liquefaction that is attributable to the imposition of stringent

Table 1.3. Potential improvements in plant systems[a]

Action	Application	Effect	Status
Increase solids content from 58.5% to 70%	Texaco	>6% increase in thermal efficiency; up to about 14% reduction in product cost	Satisfactory feeding of 70% slurry reported in literature
Decrease Texaco gasifier heat loss from 2.0% to 1.0%	Texaco	About 1% increase in plant thermal efficiency; >2% reduction in production cost	Depends upon what actual gasifier heat losses are
Employ compressors with isothermal efficiency of >70% and steam turbine drivers with >77% thermal efficiency			
Air and oxygen	All	~0.6% increase in thermal efficiency	Available
Raw gas	KT	~0.8% increase in thermal efficiency	
Utilize 80% of heating value of unburned carbon	KT and TX	Increase thermal efficiency by ~2% and decrease product cost by >0.2%	Not standard practice; design required
Use tubular reactors for methanol synthesis and conversion	All	2 to 3% improvement in thermal efficiency; cost change not determined	Available
Recover heat from process to heat boiler feedwater from condenser temperature to ~93°C (~200°F) in lieu of using 172-kPa (gage) (25-psig) steam	All	~0.1% improvement in thermal efficiency; cost change not known	Available

[a]Changes in thermal efficiencies are reported as numerical or absolute values, whereas changes in product price are relative percentages.

environmental control regulations. The reduction in plant capital and operating costs that would accrue from eliminating the stringent environmental control measures assumed in the base plant designs was estimated to be in the range of 5 to 8%, depending on the location of the plant.

1.5 CONCLUSIONS

Among the current-generation gasifiers examined (Lurgi and Koppers-Totzek), the Lurgi is more efficient when operating on western subbituminous coal and will apparently yield lower-cost products. The margin of difference between the two gasifiers is believed to be smaller than indicated by the conceptual design and cost estimate results, principally because the architect-engineer firm preparing the conceptual designs had substantial experience with Lurgi-based plants but no first-hand experiance in designing Koppers-Totzek-based plants. The uncertainties concerning the Lurgi-based plant are in the direction that economic and thermal performance may be poorer than calculated. On the other hand, it is quite probable that the performance of the plants employing entrained-flow gasifiers can be significantly improved. There are apparently opportunities for improved energy utilization in both cases. In the case of the slurry-fed gasifier, the successful use of a more concentrated coal-water slurry than was employed as a basis for the conceptual designs is highly probable and would dramatically improve gasifier performance.

In the case of the KT-type gasifier, improvements apparently are possible via alteration or elimination of quenching prior to entry of the gasifier discharge into the heat recovery train, in addition to other modifications having lesser impacts.

On the basis of the conceptual designs, the cost of gasoline production when using either Koppers-Totzek or Texaco gasifiers in an indirect coal liquefaction facility is about the same since the indicated differences are within the accuracy of the estimates. The efficiency and cost of both gasifier complexes are subject to improvement;

however, it is believed that a Texaco gasifier-based plant, assuming satisfactory operability of the Texaco gasifier, has the potential for achieving substantially higher thermal efficiency and a lower product cost than can a Koppers-Totzek based plant (or any variation of an atmospheric pressure entrained-flow gasifier) when operating on a relatively low moisture content coal. When operating on a high moisture content coal (such as the western subbituminous coals) a slurry feed gasifier will be hard-put to compete with a well-designed atmospheric pressure, dry-feed gasifier. The dry-ash Lurgi gasifier apparently has a significant advantage over the entrained-flow gasifier when it is fed with a noncaking coal.

Use of currently available, tubular reactors for methanol synthesis and methanol conversion would result in significant improvements in plant thermal efficiency over that which would be realized by using the equipment that was selected for the conceptual designs. No determination has been made of the change in product cost that might be associated with the use of such equipment. Further improvements in synthesis system efficiency apparently are attainable.